# Appendix A

# Tradeoff Studies and Options Considered

## Appendix A-1—Milling

The three most common techniques for biomass size reduction are hammer milling, knife milling, and disk refining. The power consumption for both hammer and knife milling of wheat straw, aspen, corn cobs, and corn stover has been studied by Himmel et al. (1986). This study determined that knife milling has the lowest power consumption and is the preferred technique. Thus, knife-milled material has been used for most of SERI's experimental work (Spindler et al. 1989a, Spindler et al. 1988).

However, based on the recommendations of ABB Sprout-Bauer, Inc. (1990), a manufacturer of all three types of mills, a disk refiner was chosen for this study. Both the hammer mills and the knife mills currently manufactured have limited capacity. For example, the capacity of the largest knife mill available is 8000 lb/h; thus, 20 mills are required to handle the same capacity as four disk refiners. The cost of 20 knife mills is $3.1 MM compared to the cost of $1.48 MM for four disk refiners. However, the greater power requirements for disk refiners (128 hp-h/ton, [ABB Sprout-Bauer 1990] compared to 85 hp-h/ton for knife milling [Himmel et al. 1986]) makes the total cost (operating plus capital charges) for each option approximately equal. But disk refiners were chosen because less solids handling equipment is required to feed four disk refiners when compared to the equipment required to feed 20 knife mills. Maintenance requirements are also expected to be less severe for disk refiners (knife mills would require frequent replacement of the knife blades).

## Appendix A-2—Pretreatment

### *Introduction*

Of the many biomass feedstocks available for ethanol production, one of the most abundant and cheapest is cellulosic biomass. Cellulose, a polymer of glucose, can be broken down into glucose by enzymes and then converted to ethanol by yeast. However, hydrolysis of cellulose in raw cellulosic biomass is difficult. This has been attributed to the crystallinity of cellulose and the lignin-hemicellulose sheath that surrounds the cellulose. Thus, some form of pretreatment is necessary to disrupt the lignin-hemicellulose sheath and increase the susceptibility of the cellulose to enzymatic attack. Pretreatment can also hydrolyze hemicellulose to its individual sugar components. In the case of hardwoods and wheat straw, the hemicellulose is composed primarily of the five-carbon sugar xylose, a sugar that can also be converted to ethanol. The conversion of xylose to ethanol improves the overall economics of the cellulosic biomass-to-ethanol process (Hinman et al. 1989).

Several processes can be used for pretreating biomass including autohydrolysis steam explosion, steam explosion with an acid catalyst, dilute sulfuric acid hydrolysis, and the organosolv process. The dilute-acid process uses low concentrations of sulfuric acid at relatively low temperatures (160°C for 10 min) to achieve almost complete conversion of the hemicellulose xylans to xylose (Grohmann et al. 1986, Torget et al. 1988). However, prior to this step, particle size is reduced to nearly 1.0 mm, requiring significant amounts of energy. Both steam-explosion processes use high-pressure steam and rapid depressurization to reduce the size of the biomass particle and partially hydrolyze the hemicellulose fraction. Both require operationally complex steam-explosion guns. With autohydrolysis steam explosion, the yield of xylose is low (30%-50%) (Wright 1988). However, the yield can be improved by using a catalyst, such as $SO_2$ (Schwald et al. 1989), which is the basis for the acid-catalyzed steam-explosion process. The organosolv process uses an organic solvent to dissolve the lignin and hemicellulose fractions from the cellulose. The lignin is then precipitated from solution, leaving the xylose in the liquid stream. This process is complicated and expensive, but produces a high-quality lignin stream that could be converted to high-value products.

Because of the low xylose yields and their negative impact on ethanol production economics, a detailed analysis of autohydrolysis steam explosion was not carried out in this study. Furthermore, because of the expense and complexity of the organosolv process and because there are no current large markets for high-quality lignin, this process was also not considered further. The economics of the two remaining pretreatment options, steam explosion with an acid catalyst and dilute sulfuric acid pretreatment, were evaluated with two different feedstocks: wheat straw and aspen wood chips. These materials are representative of the performance expected for two of the most abundant categories of cellulosic biomass, herbaceous and wood energy crops.

### *Methodology*

Each of the four feedstock/pretreatment combinations is analyzed on the basis of total sugar (glucose as unconverted cellulose and xylose) produced, which could then be converted to ethanol. The economic information is summarized on the basis of total sugars, assuming 90% of the cellulose leaving the pretreatment process could be converted to glucose. However, the analysis does not consider conversion of cellulose to glucose or subsequent conversion of sugars to ethanol, but only considers the pretreatment processes. A different yield of glucose from cellulose will change the absolute cost of the processes but not the relative comparison between processes.

### *Process Description*

**Acid-Catalyzed Steam Explosion With Aspen Wood.** A flowsheet of the acid-catalyzed steam explosion process for aspen wood chips is shown in Figure A-1. The design for the steam-explosion system is adopted from a design by Stone and Webster Engineering Corp. (Stone and Webster Engineering Corp. 1985), and the design of the $SO_2$ recovery section is based on the work of several groups (Schwald et al. 1989, Mackie et al. 1985, Wayman and Parekh 1988, Brownell and Saddler 1984). Finally, the design of the lime slurry preparation system section is adopted from a study by Badger Engineers, Inc. (1984).

Aspen wood chips are stored in an open pile and delivered by front-end loaders to a screen that removes oversized material, which is sent to a rechipper. The screened material is loaded into a stainless-steel steam-explosion gun. Each stainless-steel gun is a 3.5 ft (1.07 m) diameter pipe, 14.5 ft (4.45 m) long, designed for 650 psig (4480 kPa), and sealed on each end by quick-opening, full-port plug or ball valves. Once the chips are sealed inside the gun, steam and vaporized sulfur dioxide are added, and the chips are cooked for 2 min. Sulfur dioxide added to the gun is taken as a liquid from a carbon steel storage vessel, pumped to 600 psig (4236 kPa), vaporized, and mixed with steam. After cooking, the material is blown down into a 316 stainless steel flash vessel operating at 60 psig (515 kPa). The steam-exploded wood then flows to a second 316 stainless-steel atmospheric flash tank where final cooling takes place.

Vapor from each of the flash tanks is sent to partial condensers (304 stainless-steel tubes/carbon steel shell). The condensate, still containing a small amount of $SO_2$, is collected in a condensate receiver, then pumped to a 316 stainless-steel open-steam stripping column. The $SO_2$ removed from the top of the column is combined with the uncondensed $SO_2$ from the partial condensers, compressed to a liquid, and recycled back to the liquid $SO_2$ storage tank. The water from the bottom of the column is sent to waste treatment. This extensive $SO_2$ recovery system is employed to significantly reduce any discharge to the environment.

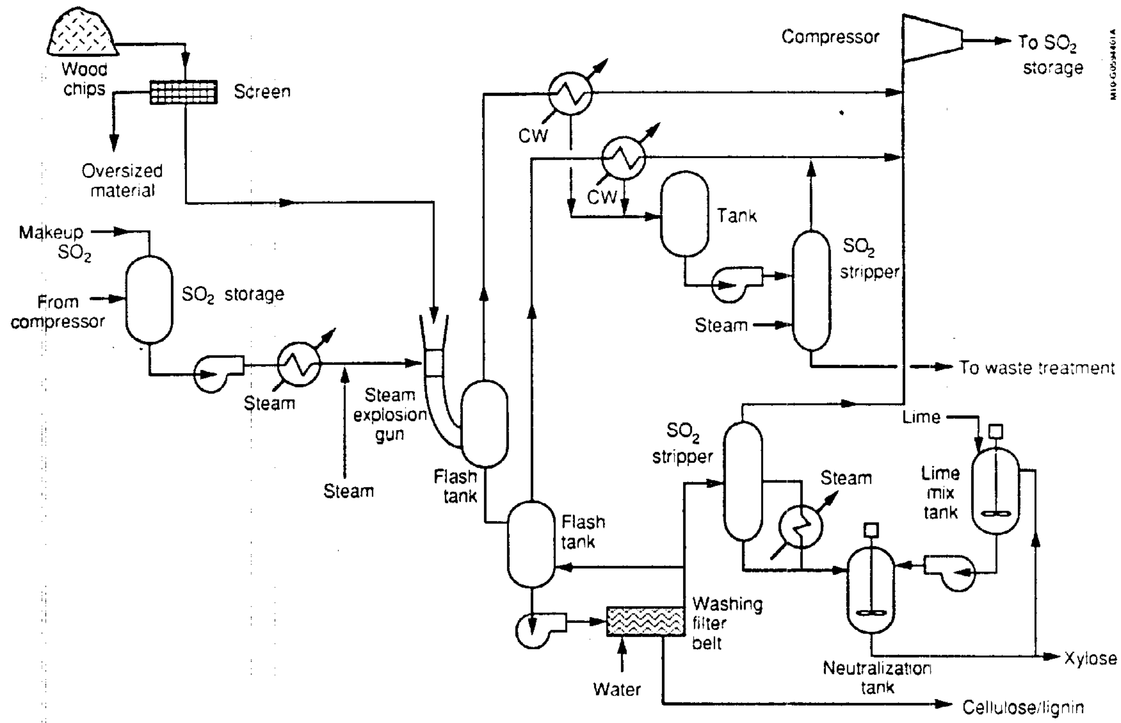



**Figure A-1. Flowsheet for acid-catalyzed steam explosion of aspen wood chips**

Exploded wood in the bottom of the atmospheric flash tank is combined with a fraction of the liquid steam from the downstream washing filter belt to obtain a pumpable slurry. The slurry is pumped to the washing filter belt to remove the $SO_2$, xylose, and other soluble material from the solid cellulose/lignin slurry. The cellulose lignin material can then be sent to an ethanol production unit. The liquid stream from the filter belt is sent to a stripping column, which is heated by a steam reboiler. The $SO_2$ removed by this column is compressed, condensed, and recycled back to the liquid $SO_2$ storage tank.

The acidified liquid (from conversion of $SO_2$ to $H_2SO_4$) from the bottom of the stripping column is sent to a 304 stainless-steel vessel for neutralization by a lime slurry, producing a neutral xylose-containing stream, which can be sent to an ethanol production unit. A small amount of this stream is taken and mixed with lime in a carbon steel vessel to produce the lime slurry.

**Acid-Catalyzed Steam Explosion With Wheat Straw.** The wheat straw plant accepts bales that are managed by a crane system and stored in a pile. After the bales are broken apart in a bale shredder, the material is screened and sent to a separator bin that removes dirt, dust, and grains from the fibers. The fibers are then sent to the steam-explosion guns. The remainder of the process is shown in Figure A-1.

**Dilute Sulfuric Acid Pretreatment With Aspen Wood.** A flowsheet for the dilute sulfuric acid process with aspen wood chips is shown in Figure A-2. The pretreatment part of this process is based on the design of Torget et al. (1988). Wood chips are screened to remove the oversize material, which is sent to a rechipper. The acceptable material is sent to a disk refiner. Here the chips are milled to reduce the particle size to approximately 1.0 mm. Following this, the milled wood enters the steam-heated impregnator where water and acid are added. This reactor ensures that acid thoroughly diffuses into the wood particles. The acidified slurry is then fed to the prehydrolysis reactor, which is steam-heated to 160°C. Both the impregnation and prehydrolysis reactors are constructed of Carpenter-20 alloy for corrosion resistance. After prehydrolysis, the wood slurry is cooled by flashing to atmospheric pressure; the slurry is subsequently conveyed to the neutralization tank. The vapors from the flash tank are condensed and sent to waste treatment. The sulfuric acid in the wood slurry is neutralized in the neutralization tank by a lime slurry and additional water is added to make a pumpable wood slurry.

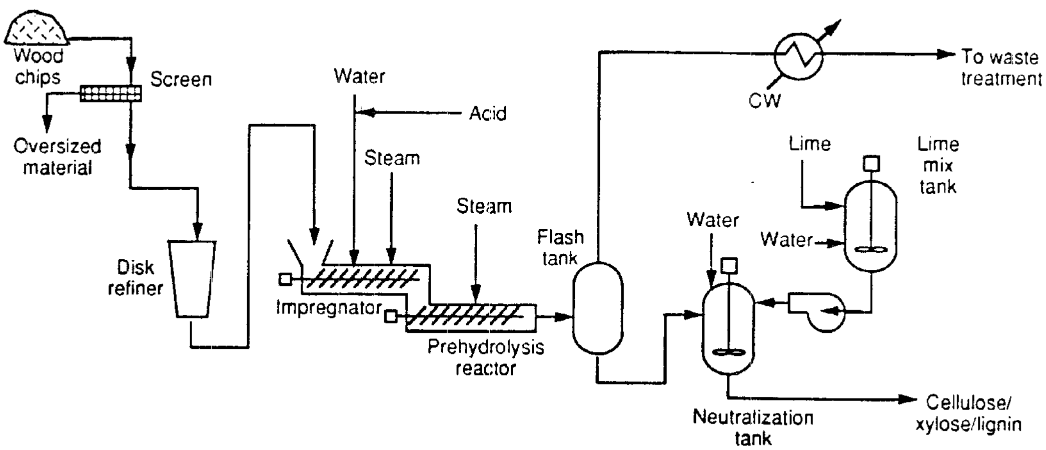



**Figure A-2. Flowsheet for dilute sulfuric acid pretreatment of aspen wood chips**

**Dilute Sulfuric Acid Pretreatment With Wheat Straw.** The design of the feed handling/storage and preparation areas is the same as that described for acid-catalyzed steam explosion with wheat straw. After screening the straw is sent to a disk refiner, and the remainder of the process is as described previously for dilute sulfuric acid pretreatment with aspen wood.

**Plant Capacity and Feed Composition.** Process flow diagrams for each feedstock were developed based on a delivered flow rate of 160,000 lb/h (72,700 kg/h) of dry feedstock to the pretreatment section of the plant. Aspen wood enters the plant at 50% moisture, and wheat straw enters the plant at 12% moisture (Strehler 1987). The chemical composition of each feedstock is given in Table A-1.

### *Design Basis*

**Acid-Catalyzed Steam Explosion.** The chips are held in the steam-explosion guns for 2.0 min at 240°C (Schwald et al. 1989). An additional 0.5 min is needed for loading and preheating, and another 0.5 min is needed for blowdown and cleaning. The total cycle time per gun is 3 min, although this has not been demonstrated at the large scale assumed for this plant.

The $SO_2$ used for steam explosion is 1.6 kg per 100 kg of dry wood, and the high-pressure steam use is 0.85 kg per kg dry wood (Schwald et al. 1989). For wheat straw, the $SO_2$ use is the same, and the high-pressure steam use is calculated to be approximately 0.45 kg per kg of dry straw because of the lower water content. The xylan conversion from a $SO_2$-catalyzed steam explosion is 75% conversion to xylose, 15% conversion to furfural, 5% unchanged, and 5% degraded to solid products. Cellulose is assumed unchanged. The $SO_2$ is converted as follows: 9.2% to sulfuric acid, 9.7% to lignin sulfonic acids, 74.2% unconverted and available for recycle, and 6.5% retained with the lignin (Schwald et al. 1989). The washing filter belt system is a five-stage washing operation, similar to a paper pulp stock washer, in which 99% of the solubles are recovered (Stone and Webster Engineering Corp. 1985).

**Dilute Sulfuric Acid Pretreatment.** The milling step requires 125 hp/ton/h (94 kW/ton) of electrical power for wood and 12.5 hp/ton/h (6.4 kW/ton) for wheat straw (ABB Sprout-Bauer 1990). The impregnator operates at 100°C with a 10-min residence time (Torget et al. 1988). The prehydrolysis

**Table A-1. Chemical Composition of Aspen Wood and Wheat Straw**

| | Aspen Wood (Badger Engineers 1984) (%) | Wheat Straw (Grohmann et al. 1986) (%) |
|---|---|---|
| Cellulose | 46.2 | 40.8 |
| Xylan | 24.0 | 27.0 |
| Lignin | 24.0 | 18.4 |
| Ash | 0.2 | 11.2 |
| Other | 5.6 | 2.6 |

reactor operates at 160°C for a 10-min residence time and with an acid concentration of 1 wt % after steam and water addition (Torget et al. 1988). Xylan is assumed converted as follows: 80% to xylose, 13% to furfural, and 7% unconverted (Grohmann et al. 1986). Also, during prehydrolysis, 4% of the cellulose is converted to glucose (Seaman 1945); the rest is assumed unchanged.

### *Capital Cost Estimate and Economic Analysis*

Heat and material balances were developed and used to specify equipment sizes. Purchased equipment cost is estimated using information from COADE (1983), Icarus Corp. (1987), Guthrie (1974), Stone and Webster Engineering Corp. (1985), and Badger Engineers, Inc. (1984). Total fixed investment is estimated as 2.85 times the purchased equipment cost (Chem Systems, Inc. 1990) plus an additional 2.0% for miscellaneous equipment. Working capital is 4.8% and startup cost is 5.0% of total fixed investment (Chem Systems, Inc. 1990). The annual capital charge (depreciation, taxes, insurance, and rate of return) is total capital invested (fixed plus working plus startup cost) times a fixed charge rate (FCR) of 0.20, typical for these types of plants (Chem Systems, Inc. 1990, Chem Systems, Inc. 1989). Chemical costs are taken from the *Chemical Marketing Reporter* (1990). Utility costs for process water, cooling water, and steam are estimated from Peters and Timmerhaus (1980), and electricity is assumed to cost $0.04/kWh. Manpower required is estimated from a previous study (Chem Systems, Inc. 1990) as 14 laborers at $29,800/y and 3 foremen at $34,000/y. Maintenance is 3.0% of total capital invested, and overhead is 65% of labor plus maintenance. Insurance and taxes are 1.5% of total fixed investment. By-product credit is taken for lignin sent to the boiler and is estimated as the heating value of lignin divided by the total heating value of the feedstock times the feedstock cost.

### *Results*

An economic summary of acid-catalyzed steam explosion for both aspen wood and wheat straw is given in Table A-2 for a feedstock cost of $42/dry ton (Wright et al. 1988) and a FCR of 0.2. The sugar selling price (glucose and xylose) as a function of feedstock cost is shown in Figure A-3 for both feedstocks. It is somewhat cheaper to produce sugars from aspen wood for the same feedstock cost. This is due to the greater amount of sugars contained in aspen wood (70% cellulose and xylans) when compared to wheat straw (68% cellulose and xylans). Also, aspen wood has a larger by-product credit because of its larger lignin content.

An economic summary of dilute sulfuric acid pretreatment for both aspen wood and wheat straw is given in Table A-3 for a feedstock cost of $42/dry ton. The sugar selling price as a function of feedstock cost is also shown in Figure A-3. In this case, aspen wood also results in a lower sugar selling price for the same feedstock cost, except below $35/dry ton where the costs are approximately the same. In this case, the advantages of the higher carbohydrate content of the aspen wood are offset by the greater electrical cost required to mill the aspen.

**Table A-2. Economy Summary of the Steam Explosion Process**

Grassroots plant, first-quarter 1990 cost
Plant Capacity: 160,000 dry lb/h
Total Capital Investment:
Aspen wood plant $53.4 MM
Wheat straw plant $54.4 MM

| | Aspen | Wheat Straw |
|---|---|---|
| | | ¢/lb sugars |
| Raw Materials | | |
| Feed | 3.28 | 3.44 |
| $SO_2$ | 0.07 | 0.08 |
| Lime | 0.01 | 0.01 |
| Utilities | | |
| Process water | 0.07 | 0.08 |
| Cooling water | 0.09 | 0.10 |
| Steam-60 psig | 0.28 | 0.26 |
| Steam-600 psig | 0.30 | 0.16 |
| Electricity | 0.03 | 0.05 |
| Labor | 0.06 | 0.06 |
| Maintenance | 0.18 | 0.19 |
| Overhead | 0.16 | 0.16 |
| Insurance and Taxes | 0.08 | 0.09 |
| By-product Credits | | |
| Lignin | 1.06 | 0.96 |
| Capital Charges | 1.30 | 1.39 |
| Totals | 4.85 | 5.11 |

Feedstock Cost: $42/dry ton
FCR: 0.20

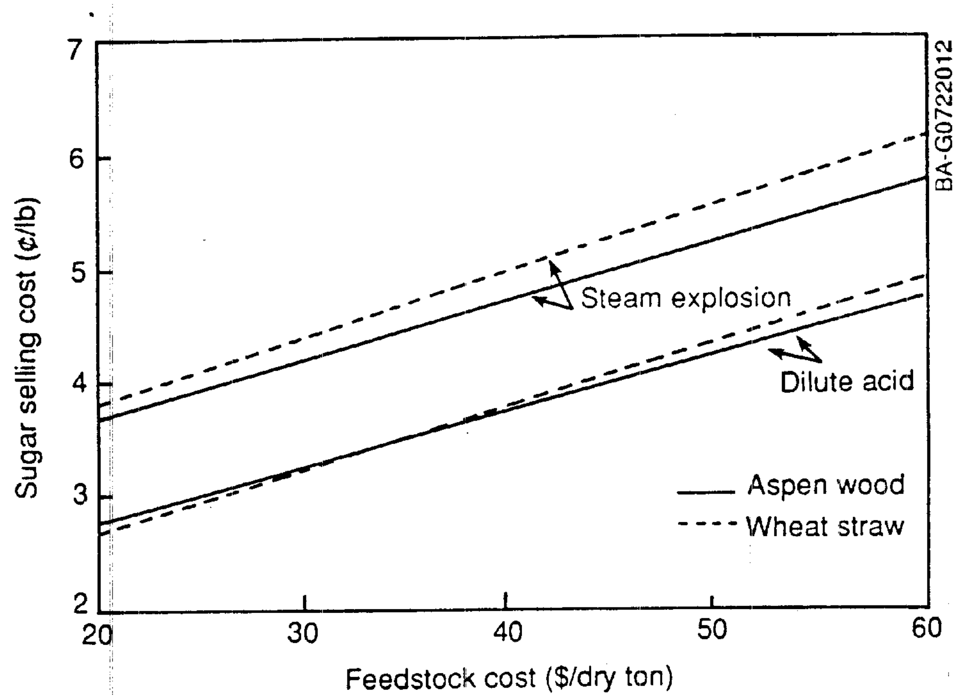



**Figure A-3. Sugar selling cost as a function of feedstock type and cost for both aspen wood and wheat straw**

### *Discussion*

The choice of feedstocks for sugar production will probably be governed by plant location. Plants situated in heavy food crop agricultural areas will probably use agricultural residues such as wheat straw. In other areas, where agricultural residues are not readily available, wood energy crops will probably be used. There is not an overwhelming advantage to either feedstock, particularly for dilute sulfuric acid, if they are obtained at the same price. However, feedstock cost does have a significant effect on the selling price of sugar. A decrease in feedstock cost from $60 to $20 per dry ton decreases the selling price by approximately 2.0 ¢/lb sugar, which is a 35% reduction for steam explosion and a 42% reduction for dilute sulfuric acid.

The data in Tables A-2 and A-3 show that dilute sulfuric acid is approximately 20% cheaper than steam explosion using the process configuration and yields assumed in this study. The higher cost for steam explosion is primarily due to the higher capital cost associated with $SO_2$ recovery. These results are based on our current understanding of each of these processes. Future improvements and changes to the process configuration could alter these results.

## Appendix A-3—Sugar Separation

In this study, prehydrolyzed wood exits the prehydrolysis reactor with a high concentration of solids (approximately 24% solids with 10% xylose). At these conditions there is no free water, and all the water, xylose, acid, and other soluble materials are absorbed into the particle. The solids concentration of a completely saturated biomass particle is 18% to 20% (Schell 1990).

**Table A-3. Economy Summary of the Dilute Acid Process**

Grassroots plant, first-quarter 1990 cost
Plant Capacity: 160,000 dry lb/h
Total Capital Investment:
Aspen wood plant $32.5 MM
Wheat straw plant $34.3 MM

| | Aspen | Wheat Straw |
|---|---|---|
| | | ¢/lb sugars |
| **Raw Materials** | | |
| Feed | 3.09 | 3.21 |
| Acid | 0.07 | 0.08 |
| Lime | 0.05 | 0.06 |
| **Utilities** | | |
| Process water | 0.03 | 0.04 |
| Cooling water | 0.02 | 0.02 |
| Steam-60 psig | 0.21 | 0.15 |
| Steam-600 psig | 0.0 | 0.0 |
| Electricity | 0.35 | 0.09 |
| Labor | 0.06 | 0.06 |
| Maintenance | 0.10 | 0.11 |
| Overhead | 0.10 | 0.11 |
| Insurance and taxes | 0.05 | 0.06 |
| **By-product credits** | | |
| Lignin | 1.00 | 0.90 |
| Capital charges | 0.75 | 0.82 |
| Totals | 3.88 | 3.91 |

Feedstock Cost:$42/dry ton
FCR: 0.20

One technique for removing xylose, acid, and other solubles from the particles is by repeated washing with water, which allows the sugars to diffuse from the particles into the bulk solution. Further recovery can then be achieved by squeezing the particles (e.g., by centrifugation), thus extracting more liquid-containing sugars. The resulting liquid is then neutralized with lime, forming calcium sulfate (gypsum). After removal of the gypsum, the stream is sent to a xylose fermentation unit for conversion of xylose to ethanol. For this option, using two centrifuges in series with a counter current flow of wash water at an assumed rate of 2.5 lb wash water per pound of solids and dewatering to 35% solids, gave a 67% recovery of xylose and diluted the xylose stream from 10.0% to 6.7%. Sugar recovery can be improved by using more wash water, but the xylose is further diluted. In order to obtain a reasonable recovery of sugars (90%), it is estimated that four or five stages of centrifugation will be required at the same wash water flow rate. Using Badger (1984) cost data for centrifuges doing similar service, the estimated purchase price for five stages of centrifugation is \$8.0 MM and total capital cost of \$22.2 MM. This is roughly 16% of our estimated total capital cost of \$141 MM. With this option, additional capital is also required for gypsum separation equipment.

Because of the large capital expense associated with this option, it is preferable to neutralize the entire stream out of prehydrolysis and send the stream to xylose fermentation. During the fermentation, xylose will diffuse out of the particle into the surrounding liquid, where it will be converted into ethanol. Because xylose is disappearing from the bulk solution (via conversion to ethanol), a concentration gradient will exist that will continue to drive xylose from the particle into the surrounding liquid. The gypsum produced by neutralization and lignin will be carried along through xylose fermentation. The advantages of this option over the previous one are reduced cost (by elimination of centrifugation and gypsum separation) and potentially higher ethanol concentration from xylose fermentation. Because of these potential advantages, this option was used for this study. However, it is important to note that no actual performance data exist for this option.

## Appendix A-4—Cellulase Production

Over the years, cellulase productivity has been improved through the development of new strains of the cellulase-producing fungus *Trichoderma reesei*. Researchers at Rutgers University developed a mutation of *T. reesei*, Rut C-30, which has significantly higher enzyme productivity than the previously used strain QM 9414. More recently, Cetus Corporation (Emeryville, CA) developed a highly productive mutation, L-27 (Shoemaker et al. 1981). In addition, experiments with various types of nutrient media have enabled cheaper ingredients, such as corn steep liquor, to be substituted for more expensive ingredients, such as proteose peptone.

Performance data for cellulase production using batch and fed-batch production techniques are available and listed in Table A-4. Fundamental kinetic information for the production of cellulase was not available. An analysis of the data shows that fed-batch production of cellulase has a higher productivity and yield than batch production. However, a true fed-batch mode of operation is not possible because the feed is wet. Fed-batch requires that essentially dry feed be added to the fermenter or that a portion of the fermenter contents be removed and slurried with the incoming feed. Because a fed-batch situation could not be envisioned, the process was designed for batch operation. Furthermore, the data are not of sufficient quality or quantity to allow any correlations to be developed. The values used for yield, substrate concentration, residence time, and specific activity are average values determined from the data for batch operation.

## Appendix A-5—Xylose Fermentation

A preliminary analysis was performed to determine the relative merits of simultaneous fermentation and isomerization of xylose (SFIX) versus *E. coli* for the fermentation of xylose to ethanol. The genetically engineered *E. coli* supplied by L. Ingram has been shown to produce high ethanol yields of 88%-95% (Spindler 1989) but requires large amounts of base to neutralize fermentation acids. SFIX does not require large amounts of base but has lower yields and requires xylose isomerase production and immobilization. Both fermentations require 2 days to achieve adequate yields.

The purpose of this analysis is to determine the best option for xylose fermentation, SFIX or *E. coli*. Initially, we will assume that xylose isomerase is infinitely stable and 100% recoverable without cost; thus, there are no capital or operating expenses associated with isomerase use. Operating and capital costs, except for base addition, are assumed to be equal for both systems, since both fermentations require 2 days. Thus, under these conditions, if the extra revenues received for the increased ethanol production

**Table A-4. Cellulase Production Data for *Trichoderma reesei*[a]**

| Cellulose Concentration (g/L) | Cell Density (g/L) | Specific Activity (IU/g eng.) | Residence Time (d) | Productivity[a] (IU/L-h) | Cellulase Yield (IU/g cellulose) | Culture[f] Method | Source |
|---|---|---|---|---|---|---|---|
| 268[b] (20)[c] | 20 | 880 | 12 | 210 | 226 | Fed-batch | Watson et al. 1984 |
| 150[b] (20)[c] | | | 12 | 105 | 203 | Fed-batch | Hendy et al. 1982 |
| 100[b] | | | 12.5 | 87 | 262 | Fed-batch | Hendy et al. 1984 |
| 80[b] (20)[a] | | | 10 | 159 | 477 | Fed-batch | McLean et al. 1985 |
| 150[b] | | | 13 | 48 | 100 | Fed-batch | Wilke, Blanch 1985 |
| 100[b] | | | 10 | 112 | 270 | Batch | Wilke, Blanch 1985 |
| 50[a] | | 1220 | 6 | 83 | 240 | Batch | Sheir-ness et al. 1984 |
| 50[a] | | 660 | 8 | 75 | 288 | Batch | Tangnu et al. 1981 |
| 50[a] | | | 4 | 68 | 130 | Batch | Hendy et al. 1982 |
| 70[a] | | 800 | 5 | 167[e] | 286 | Batch (150L) | Durand et al. 1988 |
| 50[a] | 11 | | 7 | 71[e] | 240 | Batch | Watson, Nelligan 1983 |
| 50[a] | | | 7 | 48 | 160 | Batch | Schell et al. 1990 |
| 40[a] | 14 | 100 | 2.75 | 42 | 69 | | Mohagheghi et al. 1988 |

[a] Strain Rut C-30 except for Durand et al. (1988), CL-847; and Schell et al. (1990), L-27
[b] Total cellulose added divided by fermenter volume
[c] Initial cellulose concentration
[d] Maximum productivity except as noted, calculated from cellulase yield, cellulose concentration, and residence time
[e] Average productivity, maximum not available
[f] Laboratory scale fermenters except as noted

Average productivity (fed-batch): 120
Average productivity (batch): 79
Average cellulase yield (fed-batch): 256 IU/g
Average residence time (fed-batch): 11.6 days
Average specific activity (all data): 732 IU/g
Average cellulase yield (batch): 202 IU/g
Average residence time (batch): 5.7 days

from the *E. coli* fermentation offset the base cost, this will be a more viable method for xylose fermentation.

D. Spindler's (1989) data for ethanol concentration (g/L) and NaOH usage (mL) as a function of fermentation pH are shown in Table A-5. Ethanol production from xylose fermentation (in gal) is calculated from the ethanol concentration and fermenter working volume of 2.0 L used in these fermentations. NaOH usage is calculated from the base molarity and amount used in milliliters. Because $NH_4OH$ is a cheaper base and provides some nutritional requirements, the equivalent amount of $NH_4OH$ that would be required is calculated from the ratio of molecular weights. The amount of $NH_4OH$ required per gal of ethanol produced is the ratio of $NH_4OH$ usage (in lb) to ethanol produced (gal). The base cost ($/gal ethanol produced from xylose fermentation) is $NH_4OH$ required (lb/gal ethanol) multiplied by $NH_4OH$ cost developed below.

The cost of anhydrous ammonia is $90/ton (*Chemical Marketing Reporter* 1989) ($.045/lb). Ammonia dissociates into ammonium ions according to the following reaction:

$$NH_3 + H_2O \rightarrow NH_4^+ + OH^- .$$

One lb of ammonia produces 2.06 lb of ammonium hydroxide. The cost of ammonium hydroxide is then:

$$\frac{\$0.045}{lb\ NH_3}\left(\frac{1.0\ lb\ NH_3}{2.06\ lb\ NH_4OH}\right) = \frac{\$0.0218}{lb\ NH_4OH} .$$

**Table A-5. *E. coli* Fermentation Data and Base Usage and Cost**

| ph | Ethanol Conc. (g/L) | Ethanol Prod.[a] (gal) | NaOH Usage[b] (mL) | NaOH Usage[b] (g) | $NH_4OH$ Usage (g) | $NH_4OH$ Req/. (lb/gal) | $NH_4OH$ Cost ($/gal) |
|---|---|---|---|---|---|---|---|
| 7.0 | 36.6 | .0248 | 250 | 50 | 43.7 | 3.89 | 0.84 |
| 6.5 | 39.3 | .0266 | 240 | 48 | 42.0 | 3.48 | .076 |
| 6.0 | 36.3 | .0246 | 100 | 20 | 17.5 | 1.57 | .034 |
| 5.5 | 22.6 | .0153 | 380 | 76 | 66.5 | 9.57 | .209 |

[a] 2.0 L working volume
[b] 5.0 M NaOH

The analysis continues by taking the ethanol yield for SFIX as 70% and the yield from the *E. coli* fermentation as 90%. (D. Spindler's yields for the *E. coli* fermentation were 90%, 96%, 89%, and 55% for pH controlled at 7.0, 6.5, 6.0, and 5.5, respectively.) Then, for 1 gal of ethanol produced by SFIX, 1.29 gal will be produced by the *E. coli* fermentation, and, with ethanol priced at $0.60 per gal (the cost goal of the SERI/DOE alcohol fuels program), this gives extra revenue of $0.17 for the *E. coli* fermentation. The additional cost to attain the extra revenue (at pH 7.0) is $0.11 ($0.084/gal × 1.29 gal).

Thus, the extra revenue exceeds the extra cost, and, even in light of the optimistic assumption regarding isomerase cost for the SFIX process, the *E. coli* fermentation appears to be the more economical process.

In reality, xylose isomerase is not very stable at the pHs used for xylose fermentation, as shown by Figures A-4 and A-5. These data are for the immobilization of *E. coli* xylose isomerase as reported by G. Means (1989) and coworkers at Ohio State University. The exception is the immobilization of enzyme on polyethylenimine-glutaraldehyde-silica beads (PGS), which showed no degradation after 2 days. However, even with infinite enzyme life, recovery of the immobilized enzyme from the fermentation broth in the presence of lignocellulosic particles may be difficult.

Several other problems will increase the complexity of the SFIX process. *S. pombe* (the yeast used in SFIX) requires glucose for growth, which can either be bought or obtained from hydrolysis of cellulose. The cost of glucose (cost from *Chemical Marketing Reporter* Nov. 27, 1989) per gal of ethanol is shown in Figure A-6 as a function of time to discharge of the entire fermenter contents and the cell replacement rate per fermentation (assuming cell recycle). The fermenter must be periodically dumped as the level of nonviable cells builds up. This buildup as a function of cell replacement rate per fermentation is shown in Figure A-7. Thus, low cell replacement rates and long times between fermenter dumps would be necessary to achieve reasonable glucose cost and lower enzyme replacement rates. However, this would lead to problems with buildup of lignin and unconverted cellulose in the fermenter.

Another option is xylose-fermenting yeast, but these fermentations suffer from low yields (50%-70%) and typically have longer fermentation times (2-5 days) (Skoog and Hahn-Hagerdal 1988). Another option is to buy Novo Sweetzyme Q (cost from Don Krull 1989) instead of producing xylose isomerase. Figure A-8 shows the cost of enzyme/gal ethanol out of the plant as a function of time until the fermenter

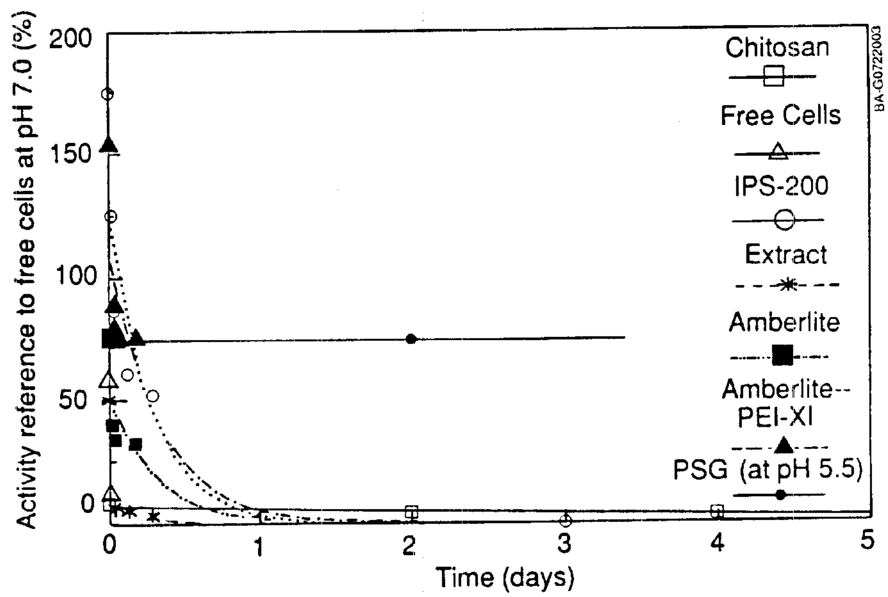



**Figure A-4. Xylose isomerase stability at pH 5.75 for different immobilization methods**

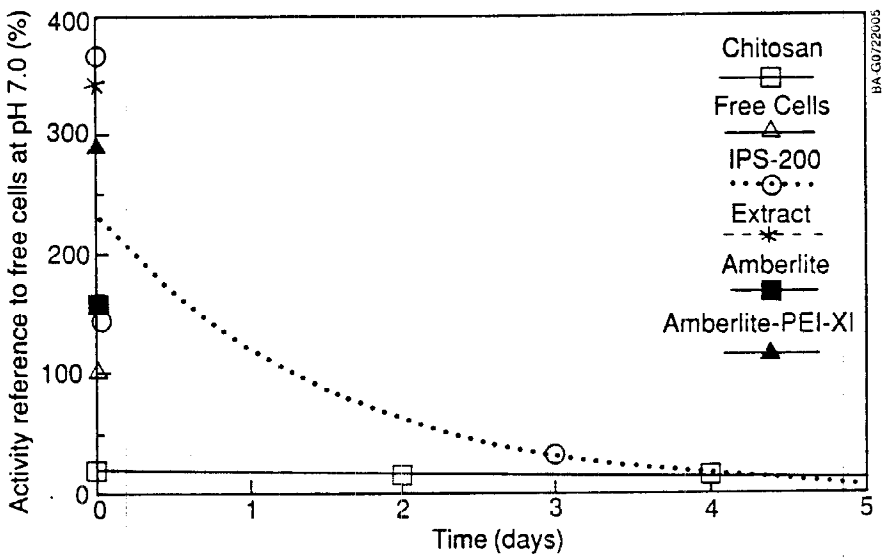



**Figure A-5. Xylose isomerase stability at pH 6.0 for different immobilization methods**

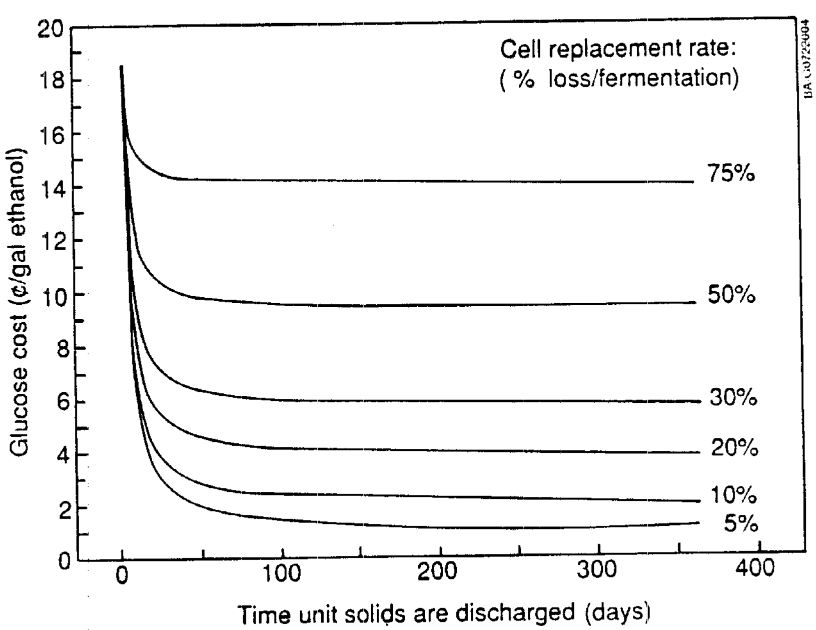



**Figure A-6. Glucose cost as a function of discharge time and cell replacement rate**

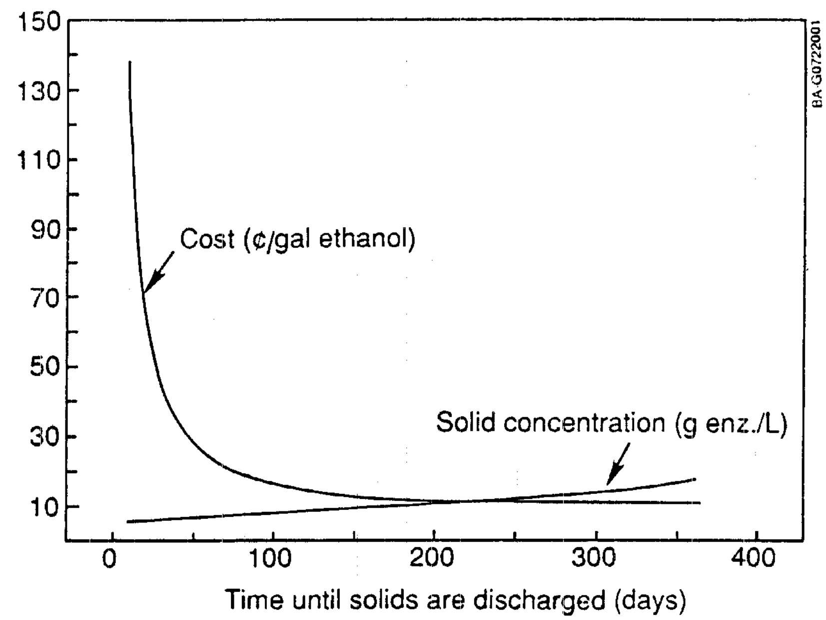



**Figure A-7. Enzyme cost and concentration as a function of discharge time**

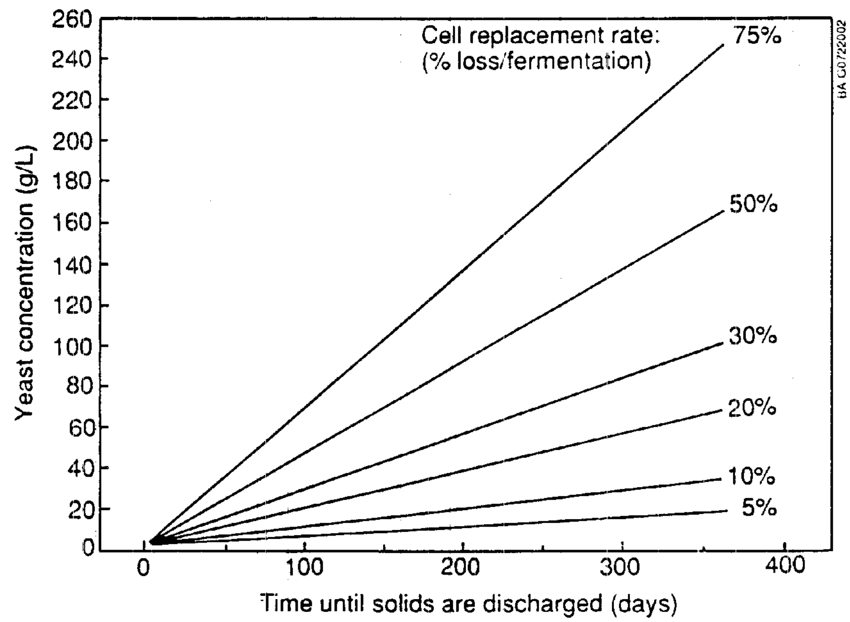



**Figure A-8. Yeast concentration as a function of discharge time and cell replacement rate**

is dumped. It is assumed that enzyme is added to maintain a constant enzyme loading and that the half-life of the enzyme is 220 days, which is estimated from Novo data. In fact, this half-life is probably much greater than the half-life that could be obtained at a pH of 5.75, which will make the true cost even higher. Even then, the minimum cost of the enzyme is approximately $0.10/gal ethanol.

For reference, SFIX performance data have been compiled in Table A-6 and xylose isomerase production data are shown in Table A-7.

## Appendix A-6—Nutrient Cost for Xylose Fermentation and SSF

Although this study assumed that all nutrient requirements are met by the recycle stream, a source of supplemental nutrients may be required. The cost for nutrients is calculated for the worst-case scenario, assuming that the recycled water does not contribute any nutrients. The nutrient requirements for both xylose fermentation and SSF are given in Table A-8. The media for xylose fermentation is an M9 minimal media with the following changes: $Na_2HPO_4$ is eliminated because buffering is not required; the concentration of $KH_2PO_4$ is doubled to supply additional phosphate; and it is assumed that only one amino acid is required at a concentration of 0.075 g/L.

The cost for nutrients is shown in Table A-9 for both xylose fermentation and SSF. The second column is the average cost for all nutrients and is determined by weighing the individual nutrient cost with the required concentration. If nutrients are required at the concentrations assumed in Table A-8, then the cost of nutrients for xylose fermentation and SSF is 21¢/gal and 14¢/gal of denatured fuel, respectively, for a total cost of 35¢/gal of denatured fuel. This is not an insignificant cost, thus, the nutrient requirements for the process are extremely important.

**Table A-6. Simultaneous Isomerization and Fermentation Data**

| Xylose Concentration (g/L) | Isomerase Loading (IU/g) | Temp. (°C) | Residence Time (d) | Ethanol Yield (%) | Cell Density (g/L) | Organism | Culture Method | Source |
|---|---|---|---|---|---|---|---|---|
| 60 | 4.2[a] | 35 | 2 | 71 | | *S. pombe* | Batch | Lastick et al. 1989 |
| 60 | 8.3[b] | 28 | 5 | 50 | 2.5 | *S. pombe* | Batch | Orton et al. 1988 |
| 120 | 10.4[b] | 35 | 2 | 63 | 45 | *S. cerevisiae* | Batch | Chiang et al. 1981 |
| 127 | | 30 | 2 | 49 | 75 | *S. cerevisiae* | Fed-Batch | Hahn-Hagerdal et al. 1986 |
| 120 | 20.8[b] | 40 | 1 | 85 | | *S. cerevisiae* | Batch | Gong et al. 1981 |
| 50 | | 29 | 12 | 9 | 1.1 | *S. pombe* | Batch | Wang et al. 1980 |

[a] Optimum rate at 2.3 IU/g
[b] Assumes 25 IU/g enzyme for Sweetzyme Q (Lastick et al. 1986)

## Table A-7. Xylose Isomerase Production Data

| Substrate | Subtrate Concentration (g/L) | Specific Activity (IU/mg) | Residence Time (h) | Isomerase Yield (IU/g) | Cell Density (g/L) | Organism | Source |
|---|---|---|---|---|---|---|---|
| Glucose | | 1.2 | 4 | | | *E. coli* | Lastic et al. 1986 |
| Glucose | 35.5[a] | | 14[b] | 40 | 15 | *E. coli* | Spencer 1989 |
| Glucose | 1.0 | .286 | | | | *E. coli* | Stevis, Ho 1985 |
| Xylose | 2.0 | .560 | | | | *E. coil* | Woycha et al. 1983 |
| Glucose | | .015 | | | | *E. coli* | Schellenburg et al. 1983 |
| Xylose | | .179 | | | | *E. coli* | Schellenburg et al. 1983 |
| Xylose/Glycerol | 20/20 | .259 | | | | *E. coli* | Schellenburg et al. 1983 |
| Xylose | 30.0 | .260 | | | | *S. violaceus* | Callens et al. 1985 |

[a] Fed-batch production
[b] First 12 h at 32°C, then temperature raised to 42°C for last 2 h.

### Table A-8. Concentration and Cost for Xylose Fermentation and SSF Nutrients

| Nutrient | Concentration (g/L) | Cost (¢/lb) |
|---|---|---|
| Xylose Fermentation | | |
| $CaCl_2$ | 0.01 | 7.65 |
| $MgSO_4$ | 0.12 | 14.00 |
| $KH_2PO_4$ | 7.00 | 6.60 |
| NaCl | 3.50 | 1.00 |
| $NH_4Cl$ | 1.00 | 18.00 |
| Amino acid | 0.075 | 1000.00 |
| SSF | | |
| $(NH)_2SO_4$ | 1.50 | 4.25 |
| $MgSO_4$ | 0.10 | 14.00 |
| $CaCl_2$ | 0.06 | 7.65 |
| Corn steep liquor | 7.50 | 11.00 |

Source for SSF nutrients: University of Arkansas

### Table A-9. Nutrient Cost

| | Average Nutrient Cost (¢/lb) | Annual Cost ($MM) | Cost (¢/gal fuel) |
|---|---|---|---|
| Xylose Fermentation | 12.35 | 12.24 | 21.1 |
| SSF | | | |
| Nutrients | 3.96 | 0.61 | 1.1 |
| Corn steep liquor | 11.00 | 7.60 | 13.1 |
| Total SSF Cost | 14.96 | 8.21 | 14.2 |
| Total Cost | | 20.45 | 35.2 |

# Appendix B

## Process Technical Data and Assumptions

### General Specifications

| | |
|---|---|
| Mixing motors | 1.0 hp/1,000 gal except as noted |
| Tank capacity | 80% full except as noted |
| Chilled water temperature | 10°C |
| Steam levels | 50 psig, 150 psig |

Feedstock Composition (dry):

| | |
|---|---|
| Cellulose | 46.2% |
| Xylan | 24.0% |
| Lignin | 24.0% |
| Ash | 0.2% |
| Others (soluble) | 5.6% |

The wood is delivered to the plant at 50% moisture.

### Feed Handling

Wood Chip Pile:

| | |
|---|---|
| Storage | 4 days |
| Losses | none (assumption) |

Mill:

| | |
|---|---|
| Particle size | 2.0-3.0 mm |
| Power requirement | 128 hp-h/dry ton (vendor number) |

### Pretreatment

Pretreatment is a two-step process. Acid impregnation is followed by prehydrolysis at a higher temperature.

Impregnation:

| | |
|---|---|
| Reactor | continuous digester (Carpenter 20 alloy) |
| Temperature | 100°C |
| Pressure | atmospheric |
| Exit solids concentration | 35 wt % (design assumption) |
| Residence time | 10 min. (assumption) |

Prehydrolysis:

| | |
|---|---|
| Reactor | continuous digester (Carpenter 20 alloy) |
| Temperature | 160°C |
| Pressure | 105 psig |
| Acid concentration | 0.85 wt % after steam addition in the prehydrolysis reactor |
| Residence time | 10 min |
| Conversions: | |
| Cellulose to glucose | 3.0% (kinetic data) |
| Cellulose to HMF | 0.1% " |
| Unconverted cellulose | 96.9% " |
| Xylan to xylose | 80.0% (experimental data) |
| Xylan to furfural | 13.0% " |
| Unconverted xylan | 7.0% " |

Flash Tank:

| | |
|---|---|
| Solids concentration | 12.0 wt % (Assumed to be a pumpable slurry) |
| Residence time | 5 min (design assumption) |
| Pressure | atmospheric |
| Mixing power | 2.0 hp/1,000 gal (assumption) |
| All furfural to overheads | (assumption) |

## Neutralization

| | |
|---|---|
| Neutralizing agent | lime |
| Residence time | 10 min (assumption) |
| Mixing power | 2.0 hp/1,000 gal (assumption) |

## Cellulase Production

Fermenters:

| | |
|---|---|
| Type | batch |
| Temperature | 28°C |
| Pressure | 10.0 psig (design assumption) |
| pH | 4.8 |
| Neutralizing agent | $NH_3$ |
| Ammonia usage | 0.045 lb/lb cellulose and xylose (experimental data) |
| Ammonia tank size | 1% of fermentation capacity (assumption) |
| Substrate | Cellulose and xylose |
| Substrate concentration | 5.0% (conforms to most experimental data) |
| Fermentation time | 5.5 days (experimental data) |
| Cycle time | 6.0 days " |
| Cellulase yield | 202 IU/g cellulose and xylose (experimental data) |
| Enzyme activity | 732 IU/g enzyme (experimental data) |
| Final cell density | 20 g/L (experimental data) |
| Specific growth rate | 0.0415/h " |
| $O_2$ uptake Rate | 42 mM $O_2$/L-h " |
| Dissolved $O_2$ | 20% of air saturation (design assumptions) |
| Antifoam use | 1.0 ml/L of fermenter volume (assumption) |

| | |
|---|---|
| Nutrients: | (from literature reference) |
| Ammonium sulfate | 1.4 g/L |
| Potassium phosphate | 2.0 g/L |
| Magnesium sulfate*$7H_2O$ | 0.3 g/L |
| Calcium chloride*$2H_2O$ | 0.4 g/L |
| Tween 80 | 0.2 g/L |
| Corn steep liquor | 15.0 g/L |

| | |
|---|---|
| Seed Fermenters (same as above except): | |
| Fermentation time | 3.5 days (from literature reference) |
| Cycle time | 4.0 days " |
| Pressure | atmospheric |
| Substrate concentration | 1.0% cellulose and xylose |
| Cell yield | 0.5 g cells/g substrate (assumption) |
| Final seed volume | 5.0% of fermenter volume (assumption) |
| Airflow | 0.2 vvm (assumption) |
| Mixing power | |
| First seed vessel | 0.5 hp/1,000 gal (assumption) |
| Other seed vessels | 1.0 hp/1,000 gal " |

| | |
|---|---|
| Sterile Feed Tank: | |
| Mixing power | maximum - 2.0 hp/1,000 gal (assumption)<br>average - one-half maximum |

| | |
|---|---|
| Cellulase Hold Tank: | |
| Mixing power | maximum - same as cellulase fermenters<br>average - one-half maximum |

## Xylose Fermentation

| | |
|---|---|
| Fermenters: | |
| Type | continuous stirred tanks in series |
| Temperature | 37°C |
| Pressure | atmospheric |
| pH | 7.0 |
| Neutralizing agent | $NH_3$ |
| Ammonia use | 0.2878 lb/lb ethanol produced (experimental data) |
| Ethanol yield | 85.5% (experimental data and assumption of 90% recovery of xylose from the particles) |
| Mixing power | 0.1 hp/1,000 gal (assumption) |
| Tank fill | 95% |
| Fermentation time | 2 days (experimental data) |
| Nutrients | none required (assumed contained in recycle water) |
| Ethanol to vent | Aspen simulation (82% recovery) |

Seed Fermenters (same as above except):

| | |
|---|---|
| Type | batch |
| Fermentation time | 12 h |
| Cycle time | 1 day |
| Substrate | xylose and glucose |
| Substrate concentration | 2.0% |
| Cell yield | 0.5 g cells/g xylose (assumption) |
| Final seed volume | 10.0% of fermenter volume (design assumption) |
| Airflow | 0.2 vvm (assumption) |

Seed Hold Tank:

| | |
|---|---|
| Mixing power | maximum - 0.1 hp/1,000 gal (assumption)<br>average - one-half maximum |

### Cellulose Fermentation

Fermenters:

| | |
|---|---|
| Type | continuous stirred tanks in series, |
| Temperature | 37°C |
| Pressure | atmospheric |
| Fermentation time | 7 days (experimental data) |
| Tank fill | 95% |
| pH | Uncontrolled |
| Cell yield | 0.5 g cells/g cellulose (assumption) |
| Mixing power | 0.1 hp/1,000 gal (assumption) |

Conversions:

| | |
|---|---|
| Cellulose to ethanol | 72.0% (experimental data) |
| Cellulose to fusel oils | 0.1% (assumption) |
| Cellulose to glycerol/ acetaldehyde | 4.9% (assumption) |
| Cellulose to cells | 10.0% " |
| Xylan to xylose | 80.0% " |
| Nutrients | none required (assumed contained in recycle water) |
| Enzyme loading | 7 IU/g cellulose |
| Ethanol tolerance | 4.5% |
| Ethanol to vent | Aspen simulation (82% recovery) |

Seed Fermenters:

| | |
|---|---|
| Same as above except | |
| Type | batch |
| Fermentation time | |
| *S. cerevisiae* | 1 day |
| *B. clausenii* | 2 day |
| Cycle time | |
| *S. cerevisiae* | 1.5 days |
| *B. clausenii* | 2.5 days |

| | |
|---|---|
| Substrate - Initial | glucose |
| - Final seed vessel | cellulose/cellulase |
| Substrate concentration | 1.0% |
| Final seed volume | 10% of fermenter volume for each culture (design assumption) |
| Airflow | 0.2 vvm (assumption) |
| Mixing power | |
| First seed vessel | 0.5 hp/1,000 gal (assumption) |
| Other seed vessels | 1.0 hp/1,000 gal (assumption) |

Seed Hold Tanks:

| | |
|---|---|
| Mixing power | maximum -0.1 hp/1,000 gal (assumption) average - one-half maximum |

## Ethanol Purification

Distillation:

| | |
|---|---|
| Ethanol concentration (Rectification column) | 95.0 wt % |
| Water to fusel oils | 5.0 lb/lb fusel oils (Badger data) |
| Ethanol concentration (Beer column) | 40.0 wt % (Badger data) |
| Reflux ratio (beer) | 0.4 (Badger data) |
| Reflux ratio (rectification) | 1.6 (Badger data) |

Lignin Separation:

| | |
|---|---|
| Solids recovery (centrifugation) | 95% (assumption) |
| Solid concentration (centrifugation) | 50% " |

## Wastewater Treatment

Anaerobic Digestion:

| | |
|---|---|
| Organics converted | 90% (lignin unconverted) (experimental data, Rivard 1990) |
| Biogas production | 0.8 lb gas/lb organics converted, balance to cell mass (experimental data, Rivard 1990) |

Aerobic Digestion:

All remaining organics degraded except lignin

| | |
|---|---|
| Solids recovery (centrifugation) | 100% (assumption) |
| Solid concentration (centrifugation) | 50% " |

### Utilities

Boiler includes Flakt drying system:

| | |
|---|---|
| Design pressure | 1100 psig, 300°F superheat |
| Efficiency | 83.5% |

Turbogenerator:

Reduces 1100 psig steam to 150 psig and 50 psig for process use, any remaining steam is condensed

| | |
|---|---|
| Efficiency | 78.5% |

Boiler Feed Water System:

| | |
|---|---|
| Water rate to boiler | 3.0% of steam usage plus direct injection (assumption) |

Cooling Water System:

| | |
|---|---|
| Water losses | 1.3% of flow for evaporation (from literature) |
| | 0.3% of flow for windage " |
| | 2.7% of flow for blowdown " |

Sterile Air System:

| | |
|---|---|
| Air temperature | 28°C |

# Appendix C

# Process Data

## Appendix C-1—Heat Capacities

| | |
|---|---|
| Wood | 0.32 Btu/lb-°F (Wenzel 1970) |
| Sulfuric acid | 0.37 Btu/lb-°F (Himmelblau 1962) |
| Ethanol vapor | 0.40 Btu/lb-°F (Yaws 1977) |
| Lime (solid) | 0.29 Btu/lb-°F (Himmelbeau 1974) |
| Gypsum | 0.26 Btu/lb-°F (Touloukian and Buyco 1970) |
| Air | 0.25 Btu/lb-°F (McCabe and Smith 1976) |
| Corn steep liquor | 1.00 Btu/lb-°F (assumed) |
| Ethanol | 0.35 Btu/lb-°F (Touloukian and Buyco 1970) |
| Carbon dioxide | 0.21 Btu/lb-°F (McCabe and Smith 1976) |
| Corn oil | 0.51 Btu/lb-°F (Perry and Chilton 1973) |
| Water vapor | 0.45 Btu/lb-°F (Touloukian and Buyco 1970) |
| Water | 1.00 Btu/lb-°F |

## Appendix C-2—Densities

| | |
|---|---|
| All process streams were estimated as water | 62.4 lb/ft$^3$ |
| Ethanol | 48.7 lb/ft$^3$ (Weast 1972) |
| Sulfuric acid | 114.2 lb/ft$^3$ (Weast 1972) |
| Lime | 139.8 lb/ft$^3$ (Weast 1972) |
| Corn oil | 57.4 lb/ft$^3$ (Weast 1972) |

## Appendix C-3—Higher Heating Values

| | |
|---|---|
| Lignin | 11478 Btu/lb (Shafizadeh 1984) |
| Cellulose | 7464 Btu/lb (Shafizadeh 1984) |
| Methane | 23984 Btu/lb (Himmelblau 1974) |
| Ethanol | 12836 Btu/lb (Weast 1972) |
| Xylose | 6747 Btu/lb (Weast 1972) |
| Xylan | 7464 Btu/lb (assumed the same as cellulose) |
| Soluble solids | 5000 Btu/lb (assumed) |
| Cellulase | 5000 Btu/lb (assumed) |
| Glycerol | 7774 Btu/lb (Weast 1972) |
| Acetaldehyde | 12835 Btu/lb (Himmelblau 1974) |
| Methane | 23984 Btu/lb (Himmelblau 1974) |

## Appendix C-4—Latent Heat

| | |
|---|---|
| Steam (50 psig) | 912 Btu/lb (Steam tables) |
| Steam (150 psig) | 857 Btu/lb (Steam tables) |
| Ethanol (12°C) | 423 Btu/lb (Touloukian and Buyco 1970) |
| Ethanol (100°C) | 324 Btu/lb (Touloukian and Buyco 1970) |

## Appendix C-5—Heat Transfer Coefficients (Tubular Exchangers)

| | |
|---|---|
| Condensing steam-liquid | 700 Btu/°F $ft^2$-h (Perry and Chilton 1973) |
| Liquid-liquid | 225 Btu/°F $ft^2$-h (Perry and Chilton 1973) |
| Condensing vapor-gas | 100 Btu/°F $ft^2$-h (Perry and Chilton 1973) |
| Condensing vapor-liquid | 400 Btu/°F $ft^2$-h (assumed) |
| Coils (coils in agitated tank) | 100 Btu/°F $ft^2$-h (Perry and Chilton 1973) |
| Gas-liquid | 60 Btu/°F $ft^2$-h (Perry and Chilton 1973) |

## Appendix C-6—Solubilities

| | |
|---|---|
| Gypsum | 0.222 g/100 cc (100°C) (Weast 1972)<br>0.241 g/100 cc (20°C) (Weast 1972) |

# Appendix D

## Spreadsheet Model

A spreadsheet model of the biomass-to-ethanol process was developed the help perform sensitivity analyses on the conceptual process design. The model includes a complete material and energy balance, capital and operating cost estimates, and economic evaluation.

The material and energy balance includes 80 streams with up to 27 components, 6 utility summaries, and 14 chemical requirement summaries. There are approximately 100 process variables that may be manipulated in the material balance in order to carry out sensitivity analyses. The variables are listed in Table D-1.

The utility summaries generated by the material and energy balance include the following:

Electricity
Low-pressure steam
High-pressure steam
Cooling water
Chilled water
Fermentation air

Feedstock, catalysts, and chemicals summaries generated include the following:

Biomass
$H_2SO_4$
Lime
$NH_3$
Corn steep liquor
Nutrients
Antifoam
Glucose
Gasoline
Diesel
Makeup water
Solids disposal
BFW chemicals
Cooling water chemicals

The capital cost estimate is generated using capacity exponents and a base case design for which a detailed cost estimate was originally made. The plant is broken down into 17 process areas and 9 utility areas as follows:

Wood Handling
Prehydrolysis
Xylose Fermentation
    Seed fermenters
    Main fermenters
    Remaining equipment
Cellulase Production
    Seed fermenters
    Main fermenters
    Remaining equipment
Simultaneous Saccharification and Fermentation
    Seed fermenters, culture 1
    Seed fermenters, culture 2
    Main fermenters
    Remaining equipment
Ethanol Recovery
    Rectification column
    Remaining equipment
Off-site Tankage
Environmental Systems
    Wastewater treatment
    Vent system
Utilities
    BFW, steam, and condensate
    Boiler
    Process water
    Turbogenerator
    Cooling water
    Chilled water
    Fermentation air
    Auxilliary utilities

Economic analyses are done on a total-plant basis and a process-unit allocated-cost basis. The cost of ethanol production is determined on both a per-year and per-gallon basis. There are approximately 25 variables in the economic analysis section of the model. These variables include:

Capital Cost
    Exponents for cost of scaled equipment, by area
    Installation factors, by area
    Working capital
Operating Costs
    Onstream factor
    Unit costs for all feedstocks, chemicals and utilities
    Labor costs

**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process**

---

Wood Handling
- Biomass feed rate, lb/h (wet basis)
- Biomass composition, wt %, (wet basis)
- Feed temperature, °F

Prehydrolysis
- $H_2SO_4$ feed rate, lb/h
- $H_2SO_4$ temperature, °F
- Water temperature, °F
- Low-pressure steam latent heat, Btu/lb
- High-pressure steam latent heat, Btu/lb
- Prehydrolysis reactor conversions
  - Xylan to xylose, %
  - Xylan to furfural, %
  - Xylan unconverted, %
  - Cellulose to glucose, %
  - Cellulose to hydroxymethylfurfural, %
  - Cellulose unconverted, %
- Latent heat of stream from blowdown, Btu/lb
- Dilution water rate to blowdown tank, lb/h

Cellulase Production
- Fraction of hydrolyzate to cellulase production
- Fraction to seed fermenters
- Dilution water rate to seed fermenters, lb/h
- Dilution water rate to main fermenters, lb/h
- Cell mass production ratio in seed fermenters, lb/lb cellulose + xylose
- Nutrient feed rate, g/L
- Corn steep liquor rate, g/L
- Base feed rate, lb $NH_3$/lb cellulose + xylose
- Antifoam feed rate, mL/L
- Enzyme yield, IU/g cellulose + xylose
- Enzyme specific activity, IU/g enzyme
- Cell mass production ratio in main fermenters, lb/lb cellulose + xylose
- Fermentation time, days
- Fermentation air rate, vvm
- Number of seed trains operating
- Time between seed batches, h
- Seed fermentation air rate, vvm
- Agitator power for main fermenters, hp/1,000 gal
- Agitator power for seed fermenters, hp/1,000 gal

**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process** (Continued)

---

Xylose Fermentation

- Fraction of remaining hydrolyzate to seed fermenters
- Dilution water rate to seed fermenters, lb/h
- Cell mass production ratio in seed fermenters, lb/lb glucose + xylose
- Base feed rate, lb $NH_3$/lb ethanol produced
- Fermentation time, days
- Fraction of xylose available for conversion
- Fraction of available xylose converted to ethanol
- Fraction of glucose converted to ethanol
- Water in fermentation off gas
- Ethanol in fermentation off gas
- Number of seed trains operating
- Time between seed batches, h
- Seed fermentation air rate, vvm
- Agitator power for main fermenters, hp/1,000 gal
- Agitator power for seed fermenters, hp/1,000 gal

Simultaneous Saccharification and Fermentation (SSF)

- Fraction to seed fermenters
- Percent cellulose converted in seed fermenters
- Percent glucose converted in seed fermenters
- Cell mass production ratio in seed fermenters, lb/lb glucose
- Dilution water rate to SSF, lb/h
- SSF fermenter conversions
    - Cellulose to ethanol + $CO_2$, %
    - Cellulose to acetaldehyde + glycerol + $CO_2$, %
    - Cellulose to yeast + $CO_2$, %
    - Cellulose to fusel oils, %
    - Glucose conversion to products above, %
    - Xylan to xylose, %
- Fermentation time, days
- Water in fermentation off gas
- Ethanol in fermentation off gas
- Fraction of total water condensed from off gas
- Fraction of total ethanol condensed from off gas
- Number of seed trains operating
- Split between two seed trains if two organisms used
- Time between seed batches, h
- Seed fermentation air rate, vvm
- Agitator power for main fermenters, hp/1,000 gal
- Agitator power for seed fermenters, hp/1,000 gal

**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process** (Concluded)

---

Ethanol Recovery
- Percent of ethanol recovered
- Percent of acetaldehyde vented
- Percent of fusel oil recovered
- Weight percent fusel oil in fusel oil product (balance assumed to be water)
- Water to fusel oil decanter, number times fusel oil rate
- Weight percent ethanol in product from ethanol distillation
- Gasoline addition rate, lb gasoline/lb ethanol
- Fraction solids recovered in lignin centrifugation
- Weight percent solids in underflow from lignin centrifuge
- Recycled process water rate, lb/h

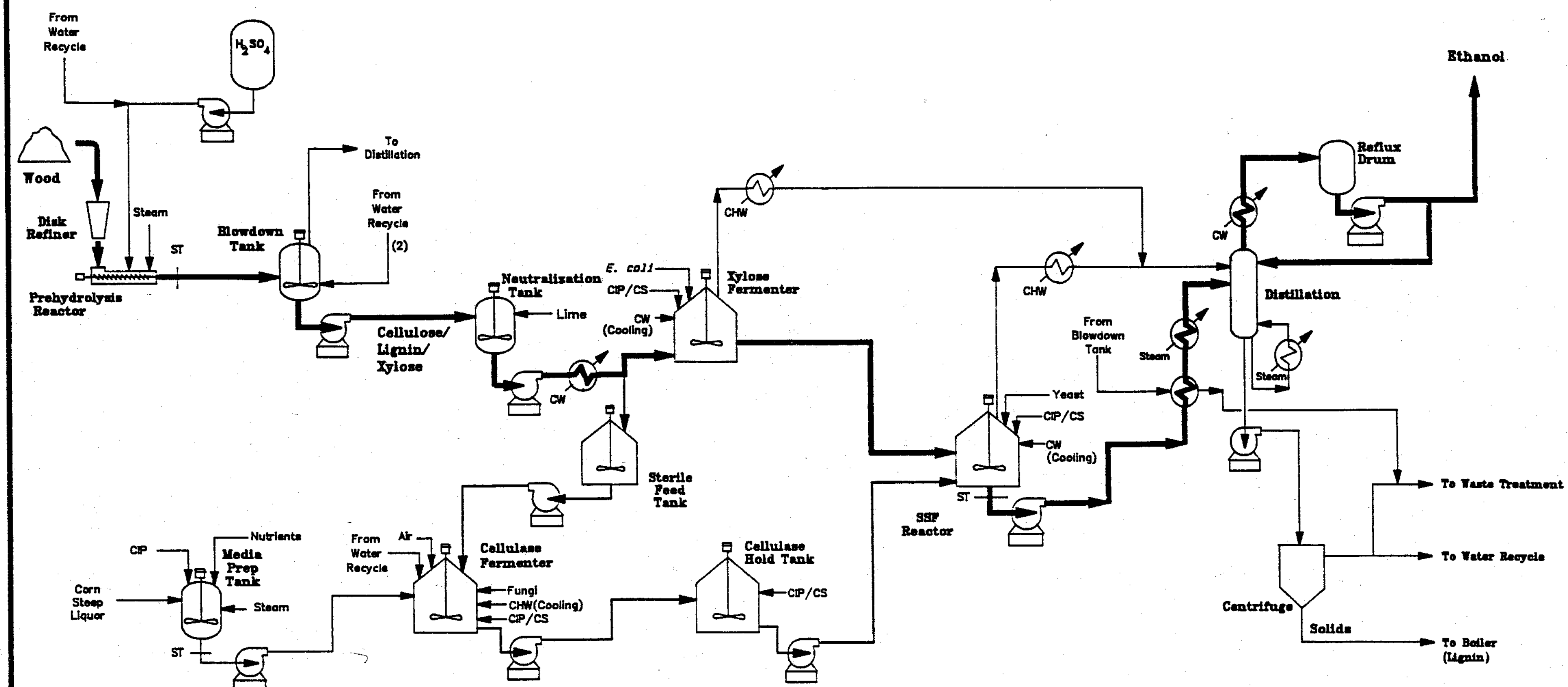
Environmental systems
- Wastewater from CIP/CS, lb/h
- Fraction of organics converted in anaerobic digestion
- Organics to biogas production ratio, lb/lb

Utilities
- Boiler efficiency, %
- Turbogenerator efficiency, %

# Appendix E

# Process Flow Diagrams

This section contains the process flow diagrams with material balances for the currently designed biomass-to-ethanol plant. Also included is a plot plan for the entire plant and a more detailed plot plan for the fermentation area.

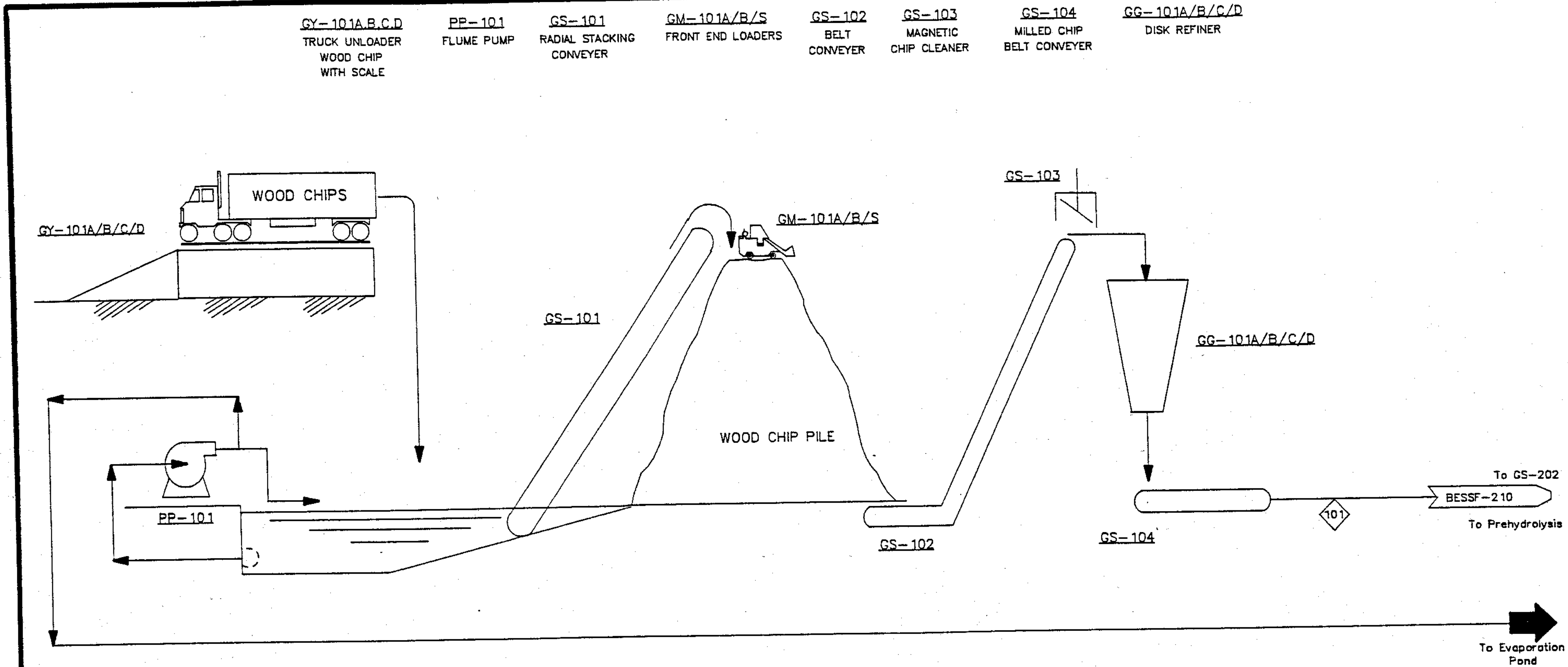



Notes

(1) ST: Sterile Boundary; CIP/CS: Clean in Place/Chemical Sterilization; CW: Cooling Water; CHW: Chilled Water

(2) Dilution water to be added if substrate concentration is such that the resulting ethanol concentration in xylose fermentation or SSF would exceed the respective ethanol tolerances or to achieve a pumpable wood slurry

| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>Process Concept | |
| Approval Date:<br>N. HINMAN | | |
| Revision Code | SERI Drawing No.: BESSF-022 | Sheet 1 of 1 |
| Misc. Data | Ref. Dwg. No. | Date 6/25/90 |

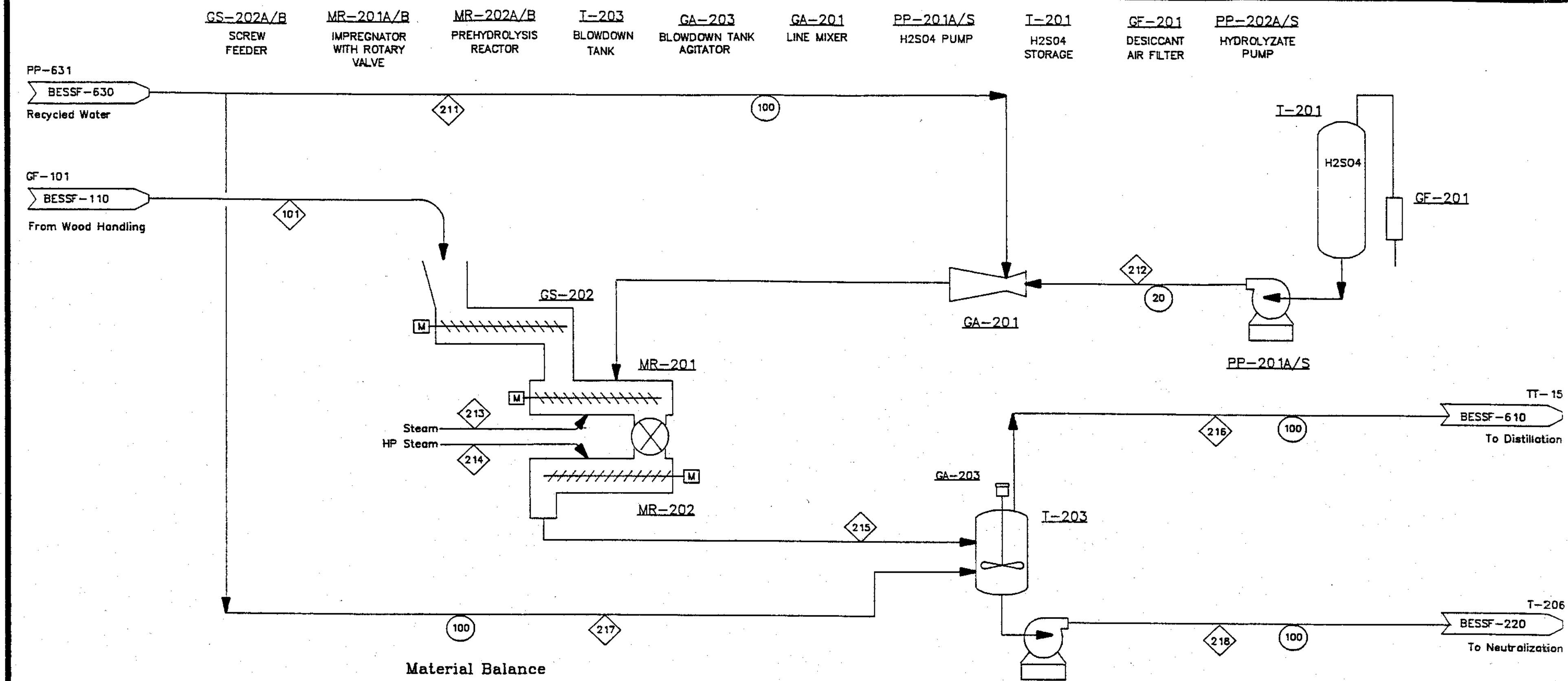



## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) | 101 |
|---|---|
| Water | 160,000 |
| Cellulose | 73,920 |
| Xylan | 38,400 |
| Soluble Solids | 8,960 |
| Ash | 320 |
| Lignin | 38,400 |
| Total | 320,000 |

◇ Stream No.

| | | | |
|---|---|---|---|
| Client<br>DOE Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | | |
| Project No<br>DOE/AFP/BESSF-10/89 | | | |
| Draftsperson Date:<br>P. WALTER | | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>WOOD HANDLING<br>RECEIVING, STORAGE, FEEDING<br>AREA 100 | | |
| Approval Date:<br>N. HINMAN | | | |
| Revision Code | SERI Drawing No.: BESSF-110 | Sheet 1 OF 1 | |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS, INC.<br>E0461-106-1 | Date 8/27/90 | |



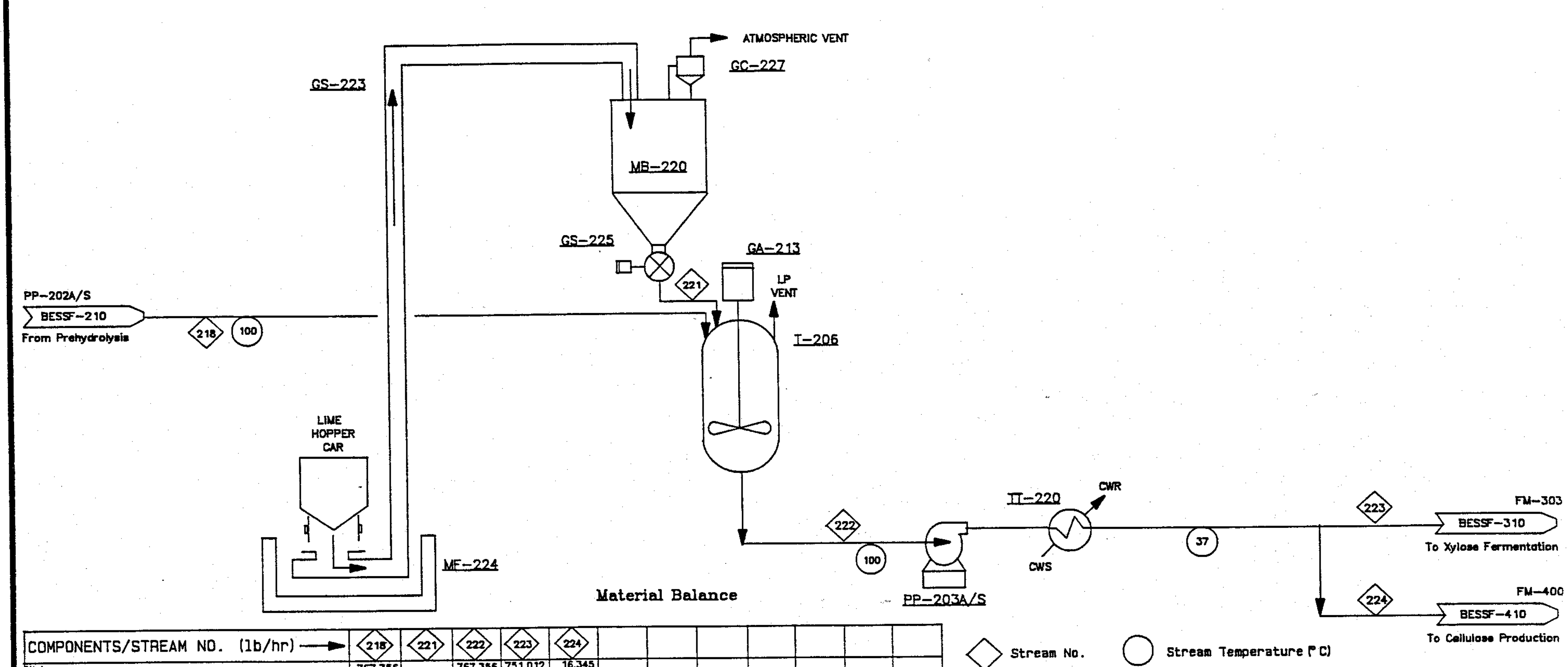

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 101 | 211 | 212 | 213 | 214 | 215 | 216 | 217 | 218 | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Water | 160,000 | 101,410 | | 30,597 | 41,354 | 330,180 | 42,148 | 479,329 | 767,356 | | | |
| Cellulose | 73,920 | 17 | | | | 71,645 | | 83 | 71,728 | | | |
| Xylan | 38,400 | 1 | | | | 2,688 | | 5 | 2,693 | | | |
| Soluble Solids | 8,960 | 1,140 | | | | 10,100 | | 5,387 | 15,487 | | | |
| Ash | 320 | 41 | | | | 361 | | 192 | 553 | | | |
| Lignin | 38,400 | 77 | | | | 38,477 | | 363 | 38,839 | | | |
| H2SO4 | | | 2,878 | | | 2,878 | | | 2,878 | | | |
| Glucose | | | | | | 2,462 | | | 2,462 | | | |
| Xylose | | 535 | | | | 35,557 | | 2,530 | 38,087 | | | |
| HMF | | 7 | | | | 65 | | 35 | 100 | | | |
| Furfural | | | | | | 3,644 | 3,644 | | | | | |
| Gypsum(Soluble) | | 129 | | | | 129 | | 611 | 740 | | | |
| Gypsum(insoluble) | | 8 | | | | 8 | | 38 | 46 | | | |
| Glycerol | | 242 | | | | 242 | | 1,145 | 1,387 | | | |
| Cell Mass | | 60 | | | | 60 | | 282 | 342 | | | |
| Total | 320,000 | 103,668 | 2,878 | 30,597 | 41,354 | 498,497 | 45,797 | 490,000 | 942,700 | | | |

◇ Stream No. ○ Stream Temperature(°C)

| | |
|---|---|
| Client<br>DOE Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P.WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>PREHYDROLYSIS<br>AREA 200 |
| Approval Date:<br>N. HINMAN | |
| Revision Code | SERI Drawing No.: BESSF-210 — Sheet 1 OF 2 |
| Misc. Data | Ref. Dwg. No.<br>Torget et al.(1988) — Date 9/25/90 |

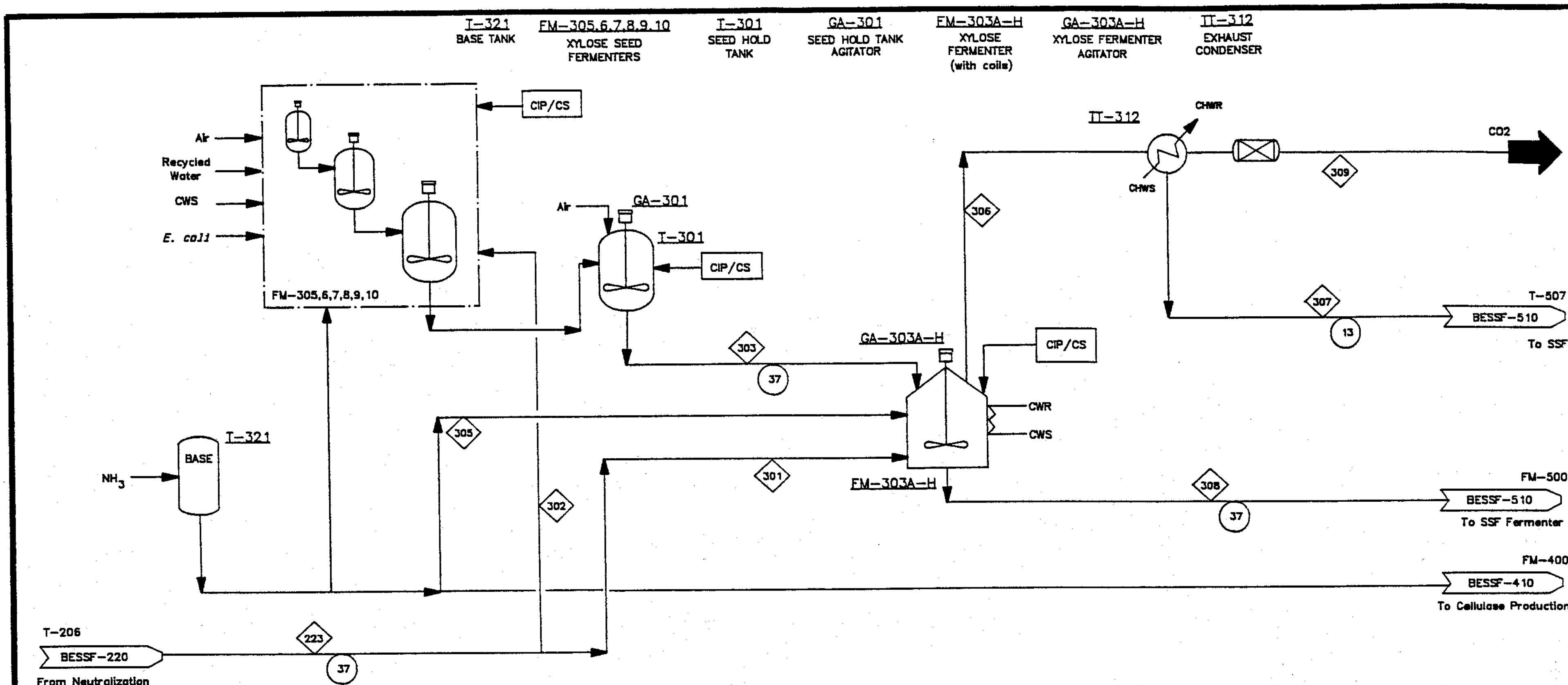
MF-224 LIME UNLOADING PIT | GS-223 LIME UNLOADING CONVEYER | MB-220 LIME STORAGE BIN | GC-227 LIME DUST CYCLONE | GS-225 LIME SOLIDS FEEDER | GA-213 NEUTRALIZATION TANK AGITATOR | T-206 NEUTRALIZATION REACTION TANK | TT-220 FEED COOLER | PP-203A/S NEUTRALIZATION HYDROLYZATE PUMP




Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 218 | 221 | 222 | 223 | 224 |
|---|---|---|---|---|---|
| Water | 767,356 | | 767,356 | 751,012 | 16,345 |
| Cellulose | 71,728 | | 71,728 | 70,200 | 1,528 |
| Xylan | 2,693 | | 2,693 | 2,636 | 57 |
| Soluble solids | 15,487 | | 15,487 | 15,157 | 330 |
| Ash | 553 | | 553 | 541 | 12 |
| Lignin | 38,839 | | 38,839 | 38,012 | 827 |
| H2SO4 | 2,878 | | | | |
| Glucose | 2,462 | | 2,462 | 2,410 | 52 |
| Xylose | 38,087 | | 38,087 | 37,276 | 811 |
| HMF | 100 | | 100 | 98 | 2 |
| Furfural | | | | | |
| Lime | | 2,128 | | | |
| Gypsum(soluble) | 740 | | 2,070 | 2,026 | 44 |
| Gypsum(insoluble) | 46 | | 3,722 | 3,643 | 79 |
| Glycerol | 1,387 | | 1,387 | 1,358 | 30 |
| Cell Mass | 342 | | 342 | 335 | 7 |
| Total | 942,700 | 2,128 | 944,828 | 924,703 | 20,125 |

| | |
|---|---|
| Client: DOE/Biofuels Program | SERI — ENGINEERING AND ANALYSIS SECTION, BIOTECHNOLOGY RESEARCH BRANCH, SOLAR FUELS RESEARCH DIVISION |
| Project No: DOE/AFP/BESSF-10/89 | |
| Draftsperson: P. WALTER Date: | |
| Designer Date: | Title: CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION — HYDROLYZATE NEUTRALIZATION — AREA 200 |
| Approval: N. HINMAN Date: | |
| Revision Code | SERI Drawing No.: BESSF-220 — Sheet 2 OF 2 |
| Misc. Data | Ref. Dwg. No. — Date 9/25/90 |



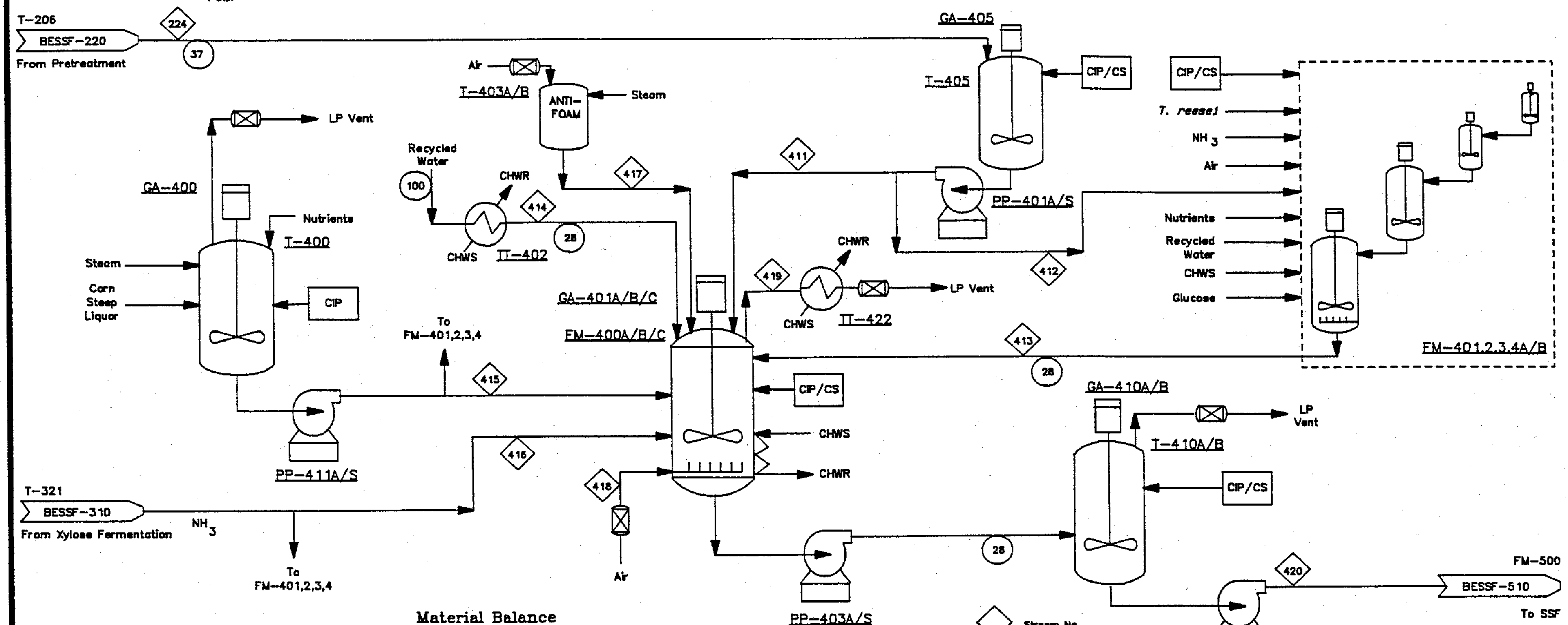

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 223 | 301 | 302 | 303 | 305 | 306 | 307 | 308 | 309 |
|---|---|---|---|---|---|---|---|---|---|
| Water | 751,012 | 719,920 | 31,092 | 89,981 | | 352 | 243 | 809,549 | 109 |
| Cellulose | 70,200 | 67,294 | 2,096 | 2,916 | | | | 70,210 | |
| Xylan | 2,636 | 2,527 | 109 | 110 | | | | 2,637 | |
| Soluble solids | 15,157 | 14,529 | 627 | 1,289 | | | | 15,819 | |
| Ash | 541 | 519 | 22 | 46 | | | | 565 | |
| Lignin | 38,012 | 36,438 | 1,574 | 1,618 | | | | 38,057 | |
| Glucose | 2,410 | 2,310 | 100 | | | | | 335 | |
| Xylose | 37,276 | 35,733 | 1,543 | | | | | 5,181 | |
| HMF | 98 | 93 | 4 | 8 | | | | 102 | |
| Gypsum(soluble) | 2,026 | 1,942 | 84 | 159 | | | | 2,101 | |
| Gypsum(insoluble) | 3,643 | 3,492 | 151 | 155 | | | | 3,648 | |
| CO2 | | | | | | 15,938 | | | 16,061 |
| NH3 | | | | | 4,285 | | | | |
| Ethanol | | | | | | 1,256 | 908 | 15,333 | 348 |
| Glycerol | 1,358 | 1,301 | 56 | 197 | | | | 1,498 | |
| Cell Mass | 335 | 321 | 14 | 1,025 | | | | 1,346 | |
| Total | 924,703 | 886,421 | 38,283 | 97,506 | 4,285 | 17,546 | 1,151 | 966,380 | 16,518 |

Stream No.  Stream Temperature(°C)

| | |
|---|---|
| Client<br>DOE/Biofuels Program | SERI |
| Project No<br>DOE/AFP/BESSF-10/89 | ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>XYLOSE FERMENTATION<br>AREA 300 |
| Approval Date:<br>N. HINMAN | |
| Revision Code | SERI Drawing No.: BESSF-310 — Sheet 1 of 1 |
| Misc. Data | Ref. Dwg. No. — Date 9/25/90 |

| T-400 | GA-400 | PP-411A/S | TT-402 | T-403A/B | FM-400A/B/C | GA-401A/B/C | TT-422 | PP-403A/S | FM-401,2,3,4A/B | T-405 | GA-405 | PP-401A/S | T-410A/B | GA-410A/B | PP-412A/S |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| MEDIA PREP TANK | PREP TANK AGITATOR | PREP TANK TRANSFER PUMP | WATER COOLER | ANITFOAM TANK | CELLULASE FERMENTER | FERMENTER AGITATOR | FERMENTER EXHAUST CONDENSER | FERMENTER RECYCLE PUMP | SEED FERMENTER PACKAGES | STERILE FEED TANK | FEED TANK AGITATOR | FEED PUMP | CELLULASE HOLD TANK | HOLD TANK AGITATOR | CELLULASE FEED PUMP |



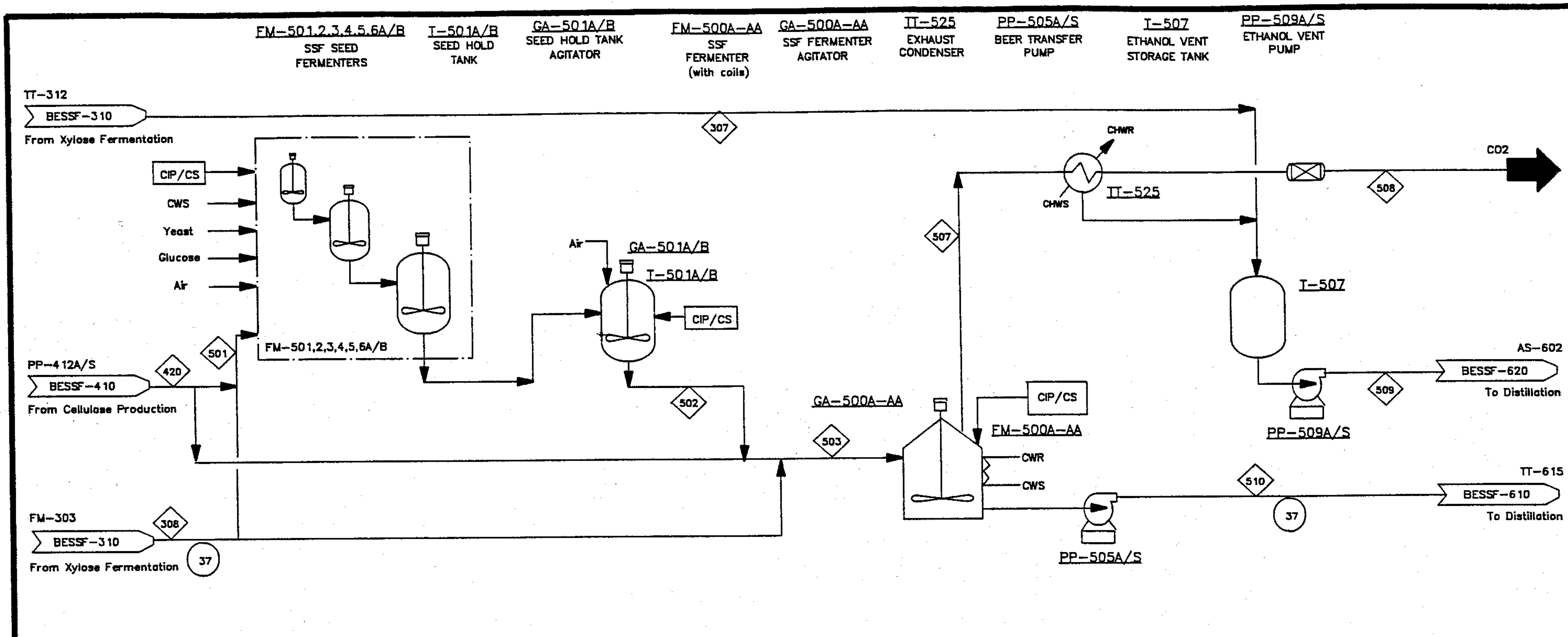

Stream No.

Stream Temperature( C)

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 224 | 411 | 412 | 413 | 414 | 415 | 416 | 417 | 418 | 419 | 420 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Water | 16,345 | 16,294 | 51 | 1,479 | 8,726 | | | | | 1,124 | 26,499 |
| Cellulose | 1,528 | 1,523 | 5 | | 2 | | | | | | |
| Xylan | 57 | 57 | | | | | | | | | 57 |
| Soluble solids | 330 | 329 | 1 | 17 | 98 | | | | | | 444 |
| Ash | 12 | 12 | | 1 | 4 | | | | | | 16 |
| Lignin | 827 | 825 | 3 | 4 | 7 | | | | | | 835 |
| Glucose | 52 | 52 | | | | | | | | | 52 |
| Xylose | 811 | 809 | 3 | | 46 | | | | | | |
| HMF | 2 | 2 | | | 1 | | | | | | 3 |
| Gypsum(Soluble) | 44 | 44 | | 2 | 11 | | | | | | 57 |
| Gypsum(insoluble) | 79 | 79 | | | 1 | | | | | | 80 |
| Air | | | | | | | | | 46,442 | 46,442 | |
| Nutrients | | | | | | 131 | | | | | |
| Cellulase | | | | | | | | | | | 657 |
| Antifoam | | | | | | | | 28 | | | |
| NH3 | | | | | | | 108 | | | | |
| Glycerol | 30 | 29 | | 4 | 21 | | | | | | 54 |
| Cell Mass | 7 | 7 | | 9 | 5 | | | | | | 616 |
| Corn Steep Liquor | | | | | | 457 | | | | | |
| Total | 20,125 | 20,062 | 62 | 1,515 | 8,920 | 588 | 108 | 28 | 46,442 | 47,565 | 29,369 |

| | | | |
|---|---|---|---|
| Client<br>DOE/Biofuels Program | | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | | |
| Draftsperson<br>P. WALTER | Date: | | |
| Designer | Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>CELLULASE PRODUCTION<br>AREA 400 | |
| Approval<br>N. HINMAN | Date: | | |
| Revision Code | | SERI Drawing No.: BESSF-410 | Sheet 1 OF 1 |
| Misc. Data | | Ref. Dwg. No. | Date 9/25/90 |



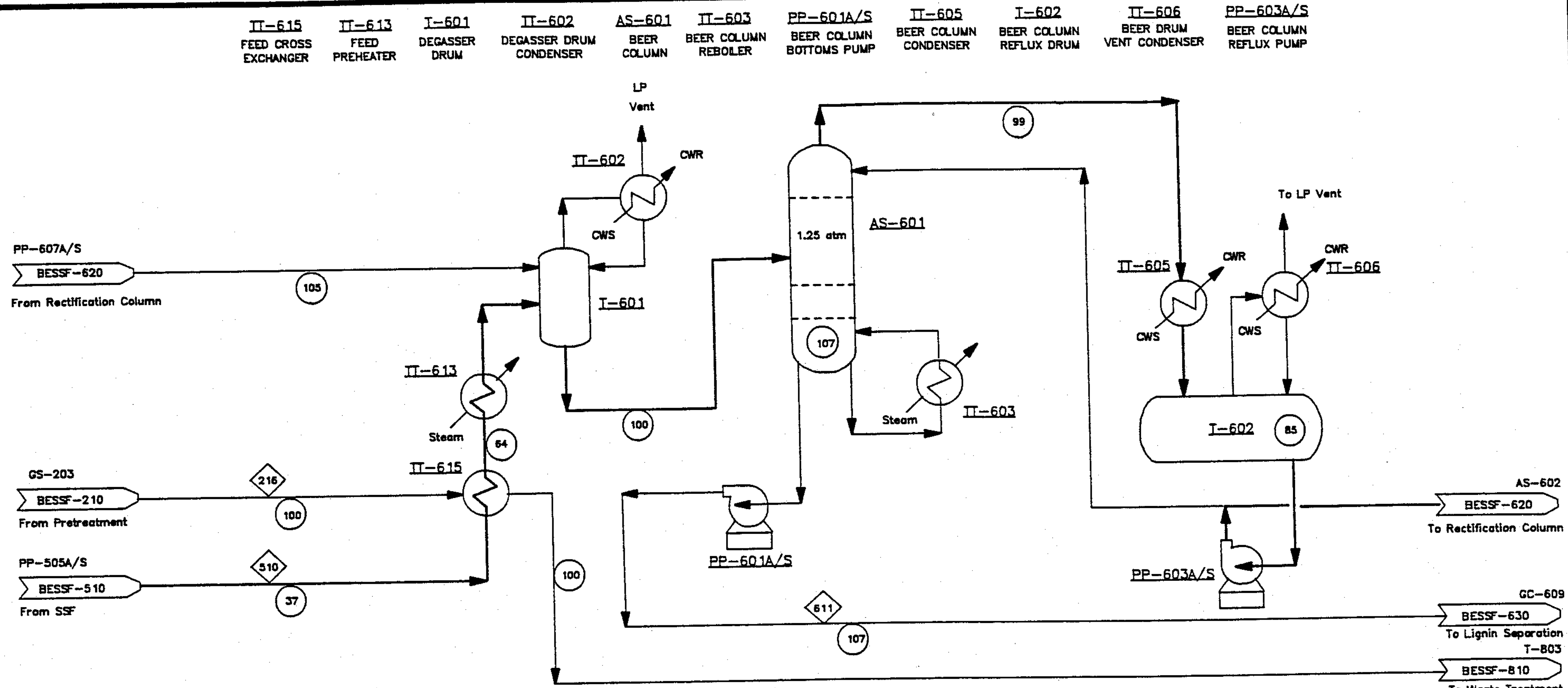

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) | 307 | 308 | 420 | 501 | 502 | 503 | 507 | 508 | 509 | 510 |
|---|---|---|---|---|---|---|---|---|---|---|
| Water | 243 | 809,549 | 26,499 | 167,209 | 167,209 | 836,047 | 655 | 163 | 735 | 832,639 |
| Cellulose | | 70,210 | | 14,042 | 11,991 | 68,159 | | | | 8,861 |
| Xylan | | 2,637 | 57 | 539 | 539 | 2,694 | | | | 539 |
| Soluble solids | | 15,819 | 444 | 3,253 | 3,253 | 16,263 | | | | 16,263 |
| Ash | | 565 | 16 | 116 | 116 | 581 | | | | 581 |
| Lignin | | 38,057 | 835 | 7,778 | 7,778 | 38,892 | | | | 38,892 |
| Glucose | | 335 | 52 | 77 | | 310 | | | | |
| Xylose | | 5,181 | | 1,036 | 1,036 | 5,181 | | | | 7,638 |
| HMF | | 102 | 3 | 21 | 21 | 105 | | | | 105 |
| Gypsum(soluble) | | 2,101 | 57 | 432 | 432 | 2,158 | | | | 2,158 |
| Gypsum(insoluble) | | 3,648 | 80 | 746 | 746 | 3,728 | | | | 3,728 |
| CO2 | | | | | | | 27,457 | 27,068 | | |
| Cellulase | | | 657 | 131 | 131 | 657 | | | | 657 |
| Ethanol | 908 | 15,333 | | 3,067 | 3,067 | 15,333 | 4,237 | 621 | 4,525 | 39,007 |
| Fusel Oils | | | | | | | | | | 76 |
| Glyceral | | 1,498 | 54 | 310 | 310 | 1,552 | | | | 3,456 |
| Acetaldehyde | | | | | | | | | | 911 |
| Cell Mass | | 1,346 | 616 | 392 | 1,570 | 3,139 | | | | 6,940 |
| | | | | | | | | | | |
| Total | 1,151 | 966,3801 | 29,369 | 199,150 | 198,198 | 994,798 | 32,349 | 27,852 | 5,259 | 962,499 |

◇ Stream No.

○ Stream Temperature(°C)

| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION | |
| Approval Date:<br>N. HINMAN | SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>AREA 500 | |
| Revision Code | SERI Drawing No.: BESSF-510 | Sheet 1 of 1 |
| Misc. Data | Ref. Dwg. No. | Date 9/25/90 |

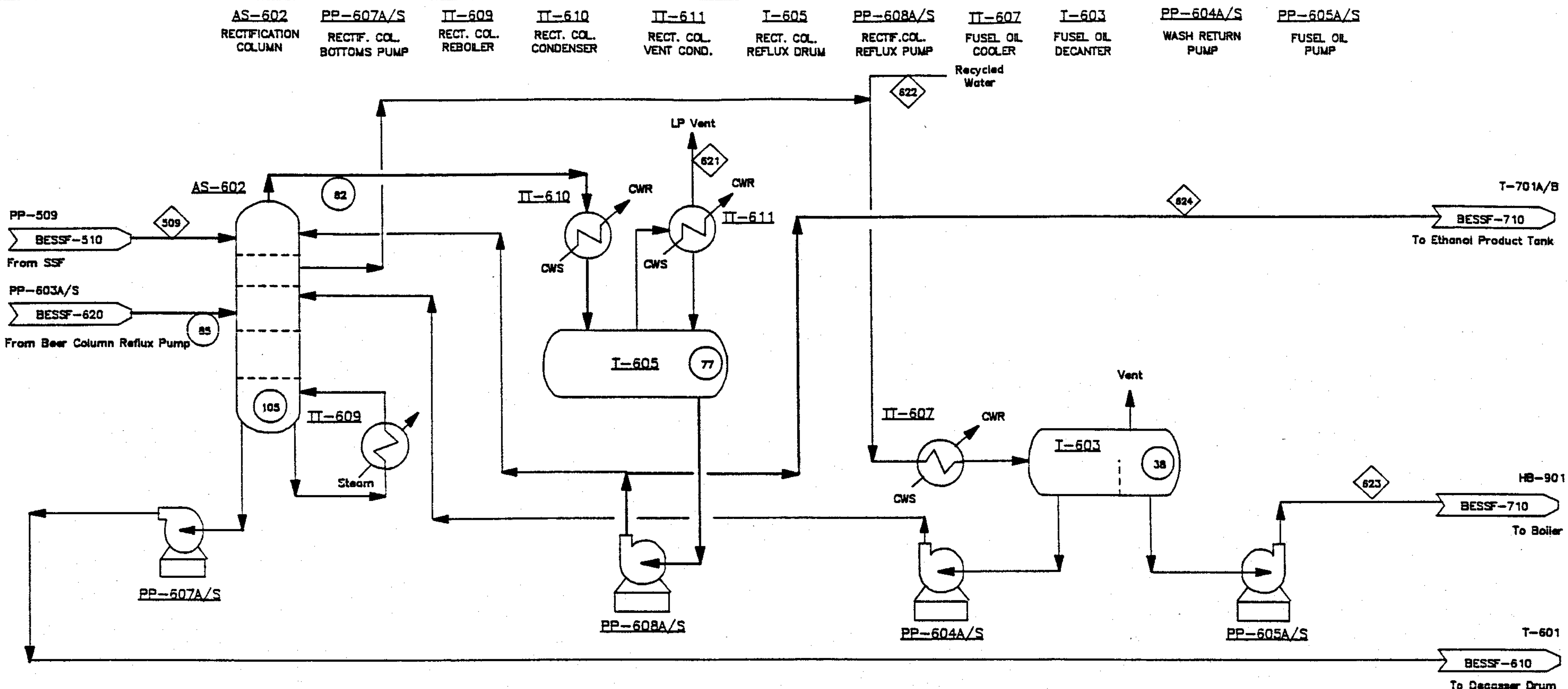



## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 216 | 510 | 611 |
|---|---|---|---|
| Water | 42,153 | 832,639 | 831,425 |
| Cellulose | | 8,861 | 8,861 |
| Xylan | | 539 | 539 |
| Soluble solids | | 16,263 | 16,267 |
| Ash | | 581 | 581 |
| Lignin | | 38,892 | 38,892 |
| Xylose | | 7,638 | 7,640 |
| HMF | | 105 | 105 |
| Furfural | 3,644 | | |
| Gypsum(soluble) | | 2,158 | 1,846 |
| Gypsum(insoluble) | | 3,728 | 4,041 |
| Cellulase | | 657 | 657 |
| Ethanol | | 39,007 | |
| Fusel Oils | | 76 | |
| Glycerol | | 3,456 | 3,457 |
| Acetaldehdye | | 911 | |
| Cell Mass | | 6,940 | 6,940 |
| | | | |
| | | | |
| Total | 45,797 | 962,499 | 921,249 |

◇ Stream No. ○ Stream Temperature(°C)

| | |
|---|---|
| Client<br>DOE/ Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>BEER DISTILLATION SYSTEM<br>AREA 600 |
| Approval Date:<br>N. HINMAN | |
| Revision Code | SERI Drawing No.: BESSF-610 — Sheet 1 OF 3 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-106-10 — Date 9/25/90 |



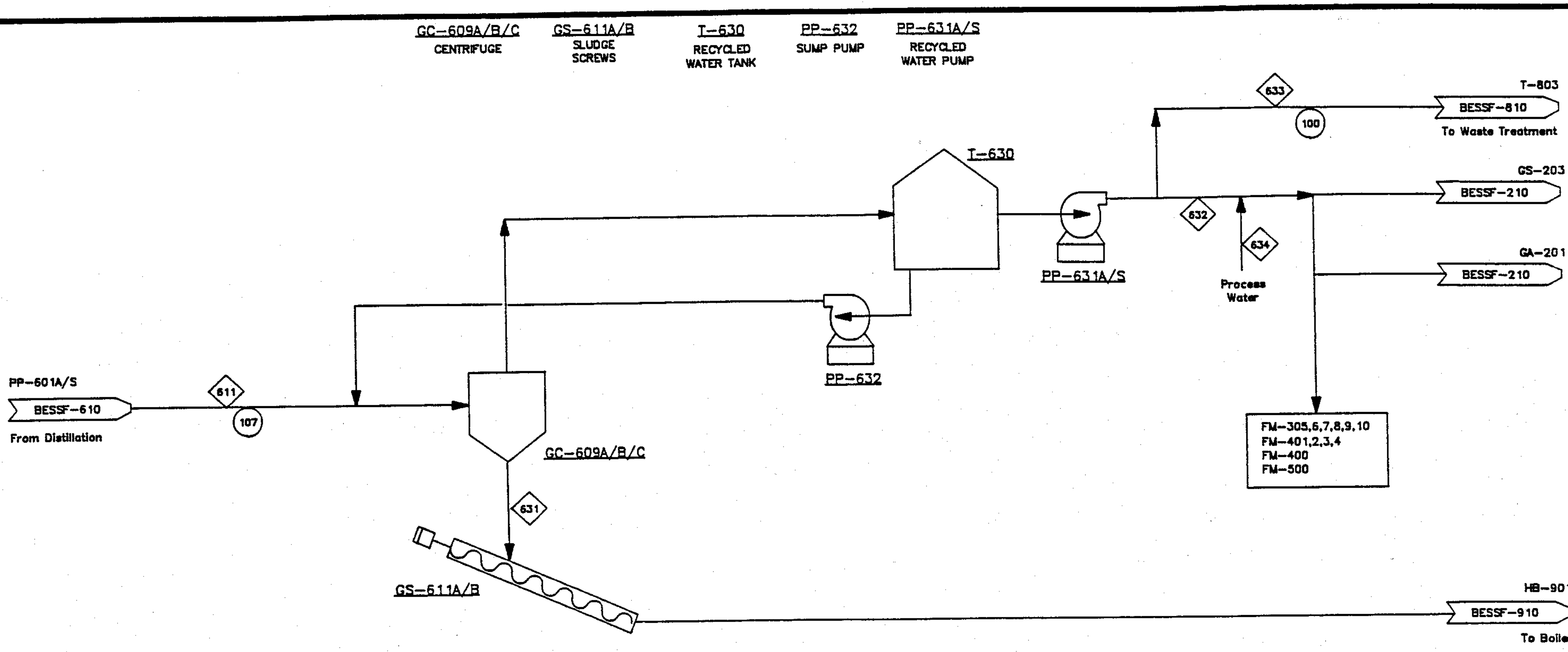

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 509 | 621 | 622 | 623 | 624 |
|---|---|---|---|---|---|
| Water | 735 | | 372 | 40 | 2,280 |
| Cellulose | | | | | |
| Xylan | | | | | |
| Soluble solids | | | 4 | | |
| Ash | | | | | |
| Lignin | | | | | |
| Xylose | | | 2 | | |
| HMF | | | | | |
| Gypsum(Soluble) | | | 1 | | |
| Gypsum(Insoluble) | | | | | |
| Cellulase | | | | | |
| Ethanol | 4,525 | 218 | | | 43,314 |
| Fusel Oils | | | | 76 | |
| Glycerol | | | 1 | | |
| Acetaldehyde | | 892 | | | 18 |
| Total | 5,259 | 1,110 | 380 | 117 | 45,612 |

◇ Stream No. ○ Stream Temperature(°C)

| | |
|---|---|
| Client: DOE/ Biofuels Program | SERI |
| Project No: DOE/AFP/BESSF-10/89 | ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Draftsperson: P WALTER Date: | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>RECTIFICATION & DEHYDRATION SYSTEM<br>AREA 600 |
| Approval: N. HINMAN Date: | |
| Revision Code | SERI Drawing No.: BESSF-620 — Sheet 2 OF 3 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-106-11 — Date 9/25/90 |

**SERI Proprietary Information**
**Do Not Copy**

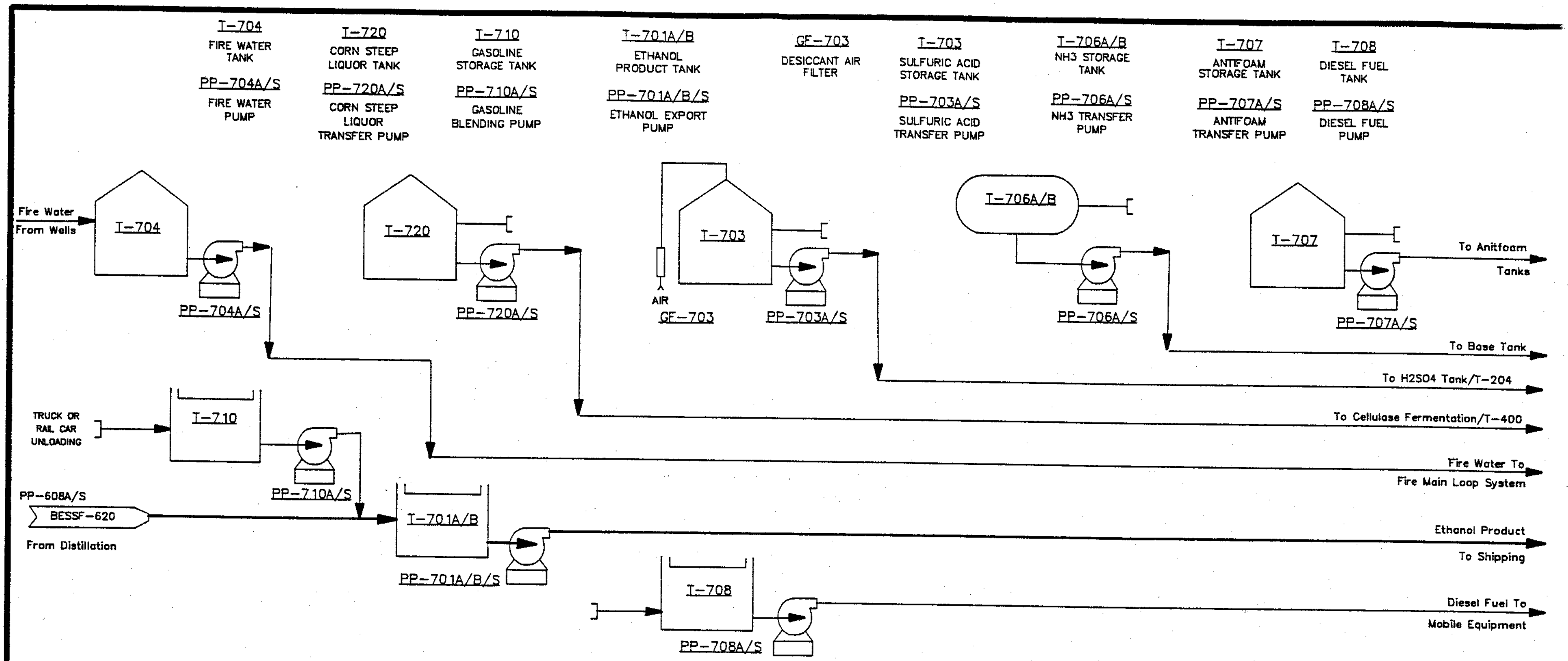



## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 611 | 631 | 632 | 633 | 634 |
|---|---|---|---|---|---|
| Water | 831,425 | 55,671 | 373,525 | 402,229 | 276,628 |
| Cellulose | 8,861 | 8,628 | 112 | 121 | |
| Xylan | 539 | 525 | 7 | 7 | |
| Soluble solids | 16,267 | 1,089 | 7,308 | 7,869 | |
| Ash | 581 | 39 | 261 | 281 | |
| Lignin | 38,892 | 37,871 | 492 | 529 | |
| Xylose | 7,640 | 512 | 3,432 | 3,696 | |
| HMF | 105 | 7 | 47 | 51 | |
| Gypsum(soluble) | 1,846 | 124 | 829 | 893 | |
| Gypsum(insoluble) | 4,041 | 3,935 | 51 | 55 | |
| Cellulase | 657 | 44 | 295 | 318 | |
| Glycerol | 3,457 | 231 | 1,553 | 1,672 | |
| Cell Mass | 6,940 | 6,758 | 88 | 94 | |
| Total | 921,249 | 115,433 | 388,000 | 417,816 | 276,628 |

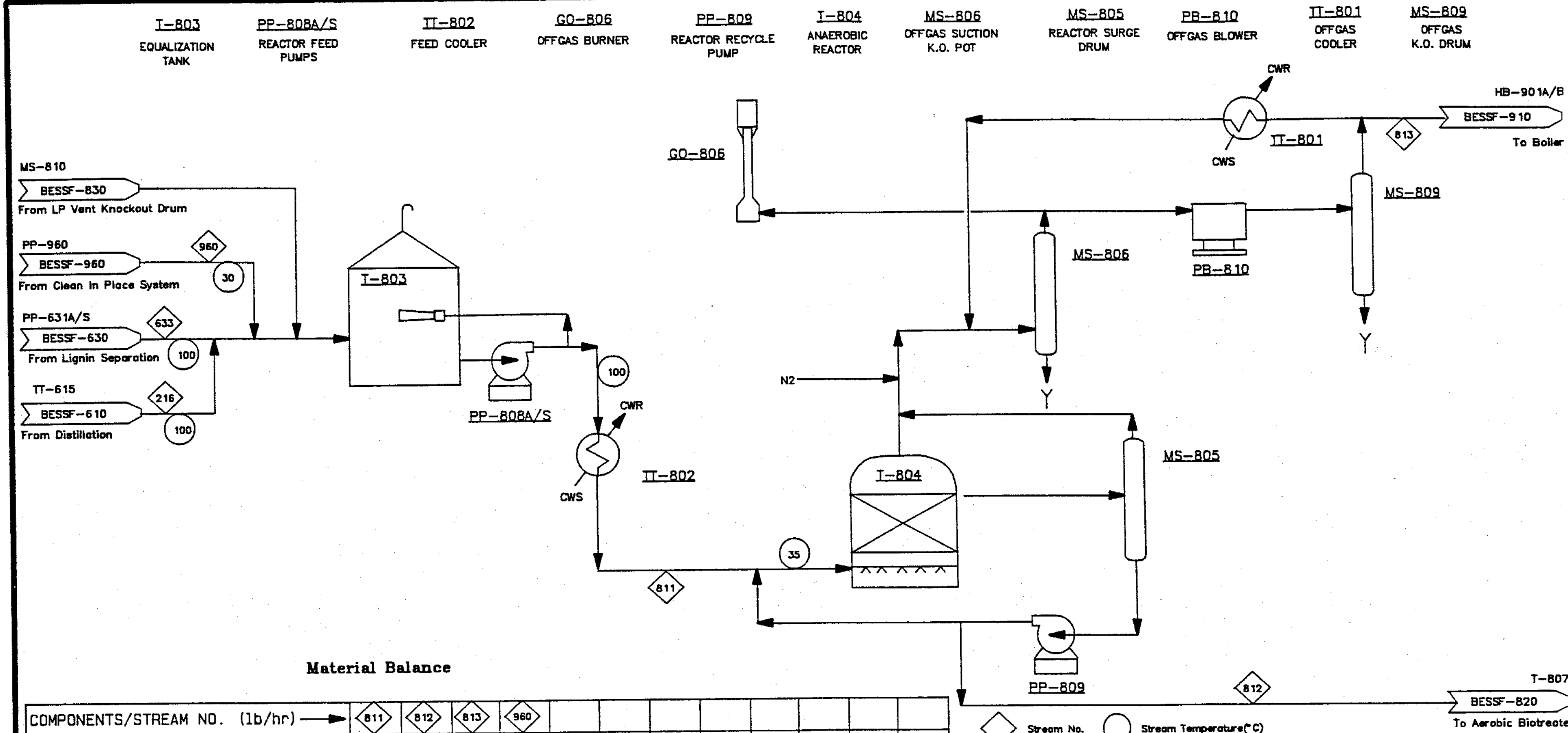
Stream No.    Stream Temperature(°C)

| | |
|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>LIGNIN SEPARATION<br>AREA 600 |
| Approval Date:<br>N. HINMAN | |
| Revision Code | SERI Drawing No.: BESSF-630 — Sheet 3 OF 3 |
| Misc. Data | Ref. Dwg. No. — Date 9/25/90 |




| | |
|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P. Walter | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>OFFSITE TANKAGE SYSTEM<br>AREA 700 |
| Approval Date:<br>N. Hinman | |
| Revision Code | SERI Drawing No.: BESSF-710 — Sheet 1 OF 1 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-13 — Date 9/25/90 |

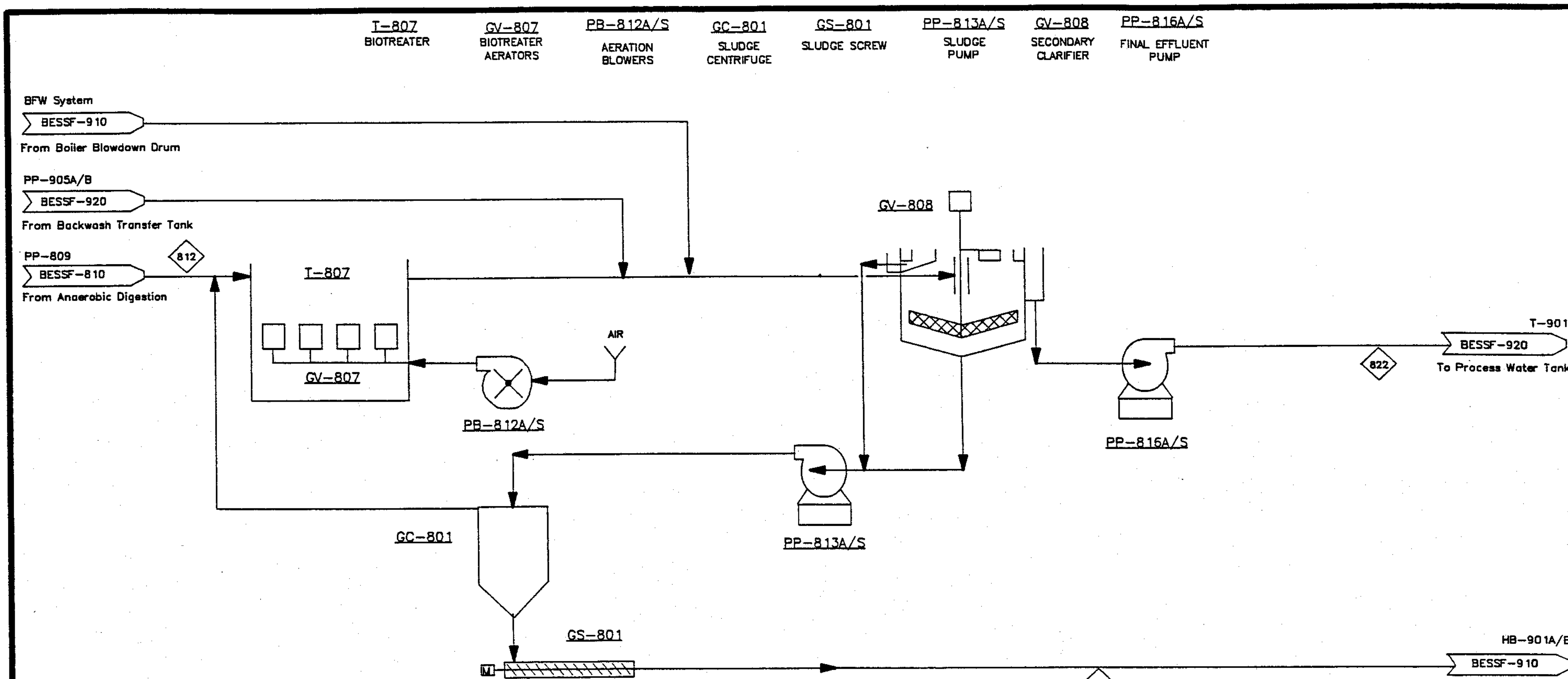
**SERI Proprietary Information**
**Do Not Copy**




## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 811 | 812 | 813 | 960 |
|---|---|---|---|---|
| Water | 444,702 | 444,242 | 460 | 320 |
| Cellulose | 121 | 12 | | |
| Xylan | 7 | 1 | | |
| Soluble solids | 7,869 | 787 | | |
| Ash | 281 | 281 | | |
| Lignin | 529 | 529 | | |
| Xylose | 3,696 | 370 | | |
| HMF | 51 | 5 | | |
| Furfural | 3,644 | 364 | | |
| Gypsum(soluble) | 893 | 893 | | |
| Gypsum(insoluble) | 55 | 55 | | |
| CO2 | | | 6,793 | |
| Cellulase | 318 | 32 | | |
| Glycerol | 1,672 | 167 | | |
| Cell Mass | 94 | 3,155 | | |
| Methane | | | 5,535 | |
| | | | | |
| | | | | |
| | | | | |
| Total | 463,933 | 450,892 | 12,789 | 320 |

| | |
|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>ENVIRONMENTAL SYSTEM<br>WASTEWATER TREATMENT<br>AREA 800 |
| Approval Date:<br>N. HINMAN | |
| Revision Code | SERI Drawing No.: BESSF-810 — Sheet 1 OF 3 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-15 — Date 9/26/90 |

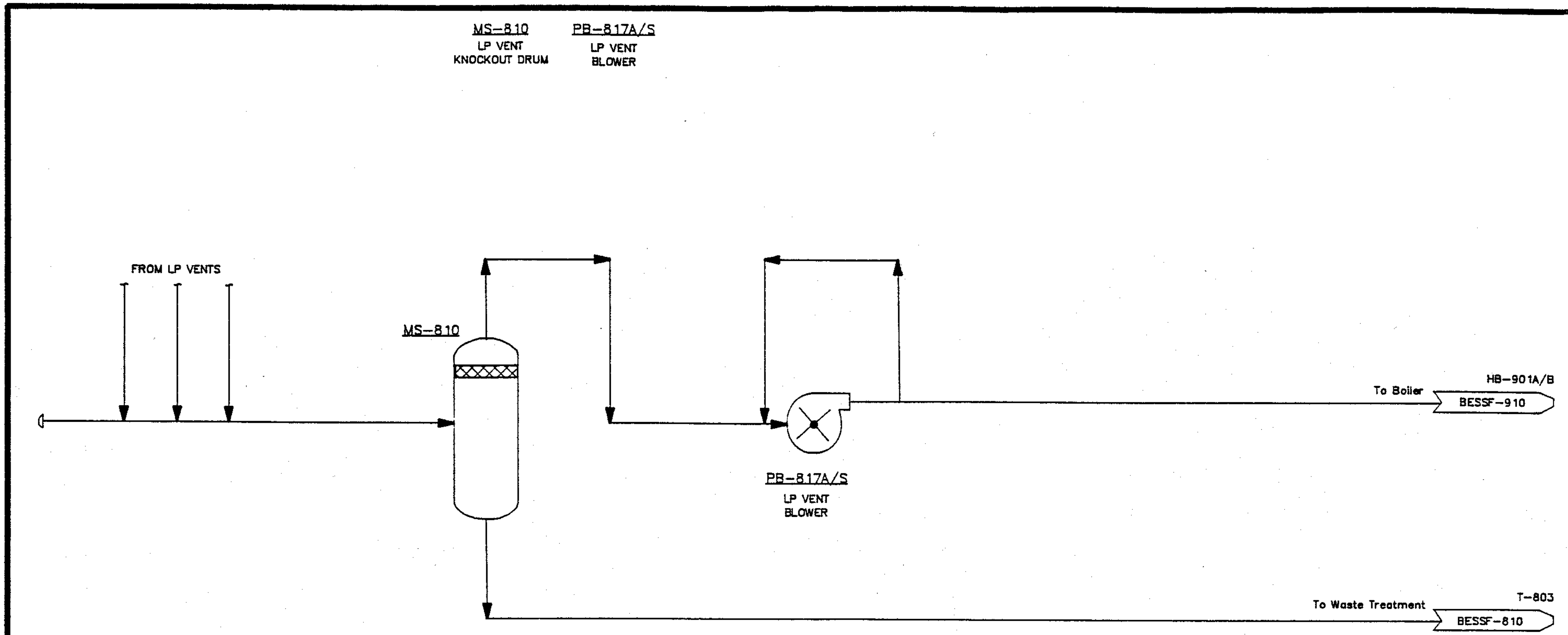



## Material Balance

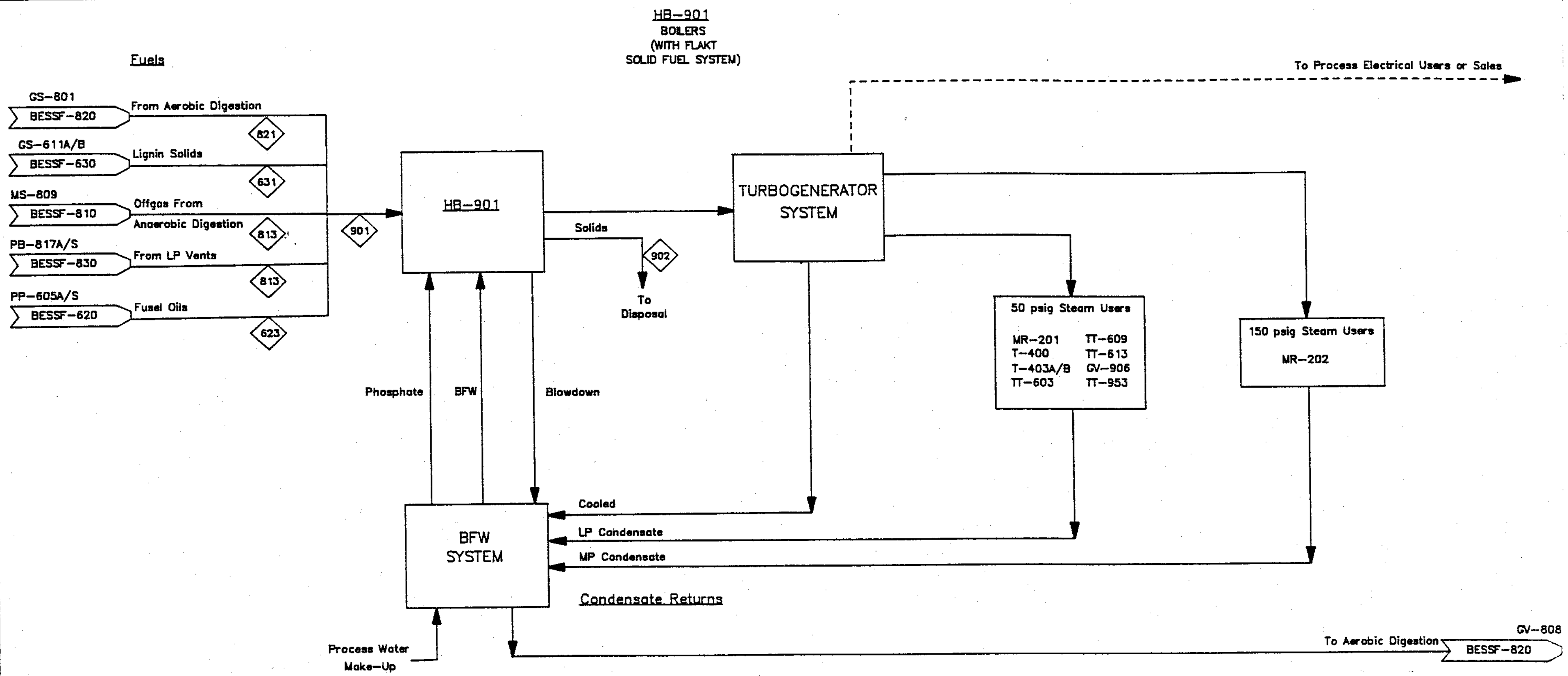
| COMPONENTS/STREAM NO. (lb/hr) → | 812 | 821 | 822 |
|---|---|---|---|
| Water | 444,242 | 4,577 | 439,665 |
| Cellulose | 12 | | |
| Xylan | 1 | | |
| Soluble solids | 787 | | |
| Ash | 281 | | 281 |
| Lignin | 529 | 529 | |
| Xylose | 370 | | |
| HMF | 5 | | |
| Furfural | 364 | | |
| Gypsum(insoluble) | 893 | 893 | |
| Gypsum(soluble) | 55 | | |
| Cellulase | 32 | | |
| Glycerol | 167 | | |
| Cell Mass | 3,155 | 3,155 | |
| Total | 450,892 | 9,154 | 439,946 |

◇ Stream No.

| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>ENVIRONMENTAL SYSTEM<br>WASTEWATER TREATMENT<br>AREA 800 | |
| Approval Date:<br>N. HINMAN | | |
| Revision Code | SERI Drawing No.: BESSF-820 | Sheet 2 OF 3 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-16 | Date 9/26/90 |




| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>ENVIRONMENTAL SYSTEM<br>PLANT VENT SYSTEM<br>AREA 800 | |
| Approval Date:<br>N. HINMAN | | |
| Revision Code | SERI Drawing No.: BESSF-830 | Sheet 3 OF 3 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC.<br>E0461-106-17 | Date 6/21/90 |

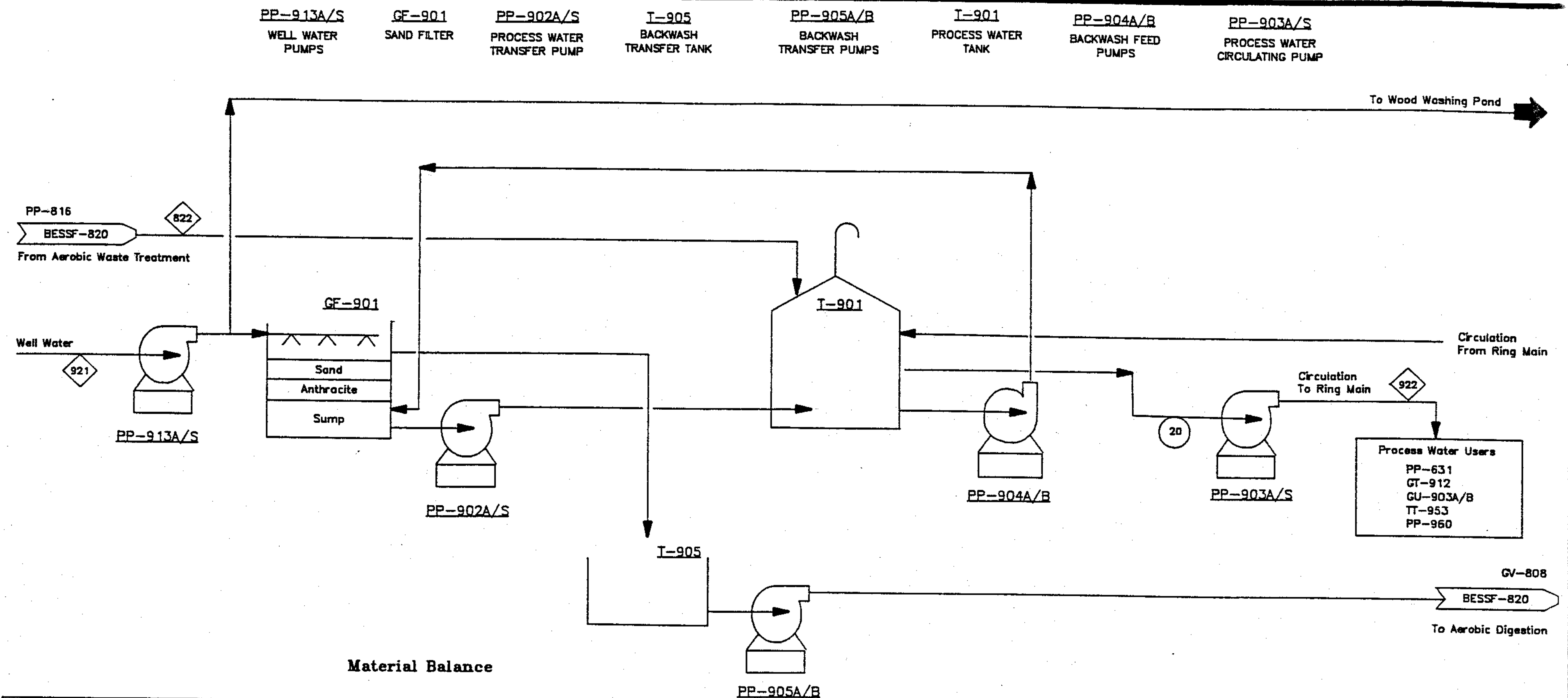



## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 901 | 902 |
|---|---|---|
| Water | 60,708 | |
| Cellulose | 8,628 | |
| Xylan | 525 | |
| Soluble solids | 1,089 | |
| Ash | 39 | 39 |
| Lignin | 38,400 | |
| Xylose | 512 | |
| HMF | 7 | |
| Gypsum(soluble) | 1,017 | 1,017 |
| Gypsum(insoluble) | 3,935 | 3,935 |
| CO2 | 6,793 | |
| Cellulase | 44 | |
| Glycerol | 231 | |
| Cell Mass | 9,913 | |
| Methane | 5,535 | |
| Fusel Oils | 75 | |
| | | |
| | | |
| | | |
| Total | 137,451 | 4,991 |

◇ Stream No.

| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Project No<br>DOE/AFP/BESSF-10/89 | | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>UTITILIES<br>BOILERS & STEAM DISTRIBUTION<br>AREA 900 | |
| Approval Date:<br>N. HINMAN | | |
| Revision Code | SERI Drawing No.: BESSF-910 | Sheet 1 OF 6 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-18 | Date 9/26/90 |

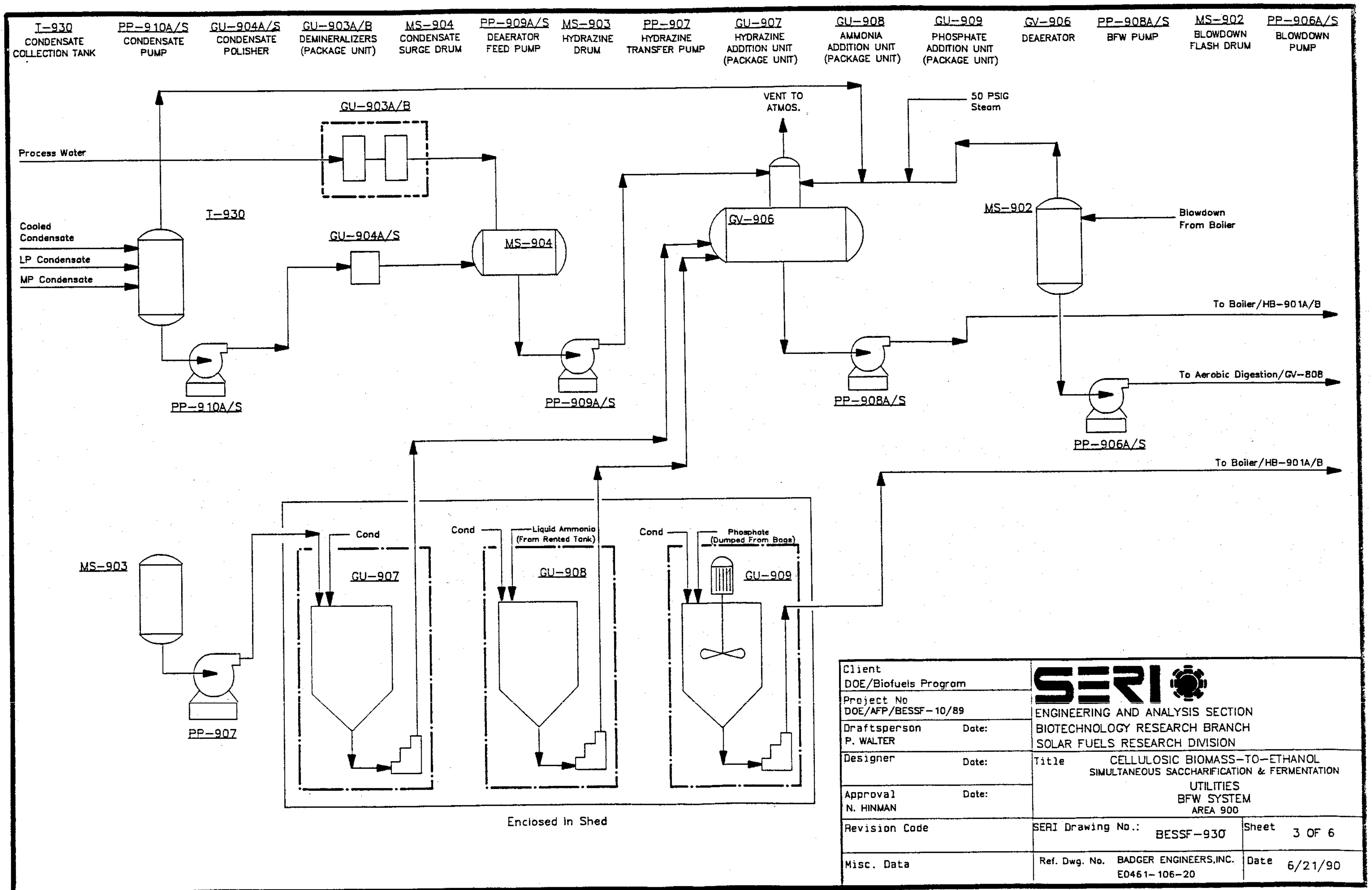


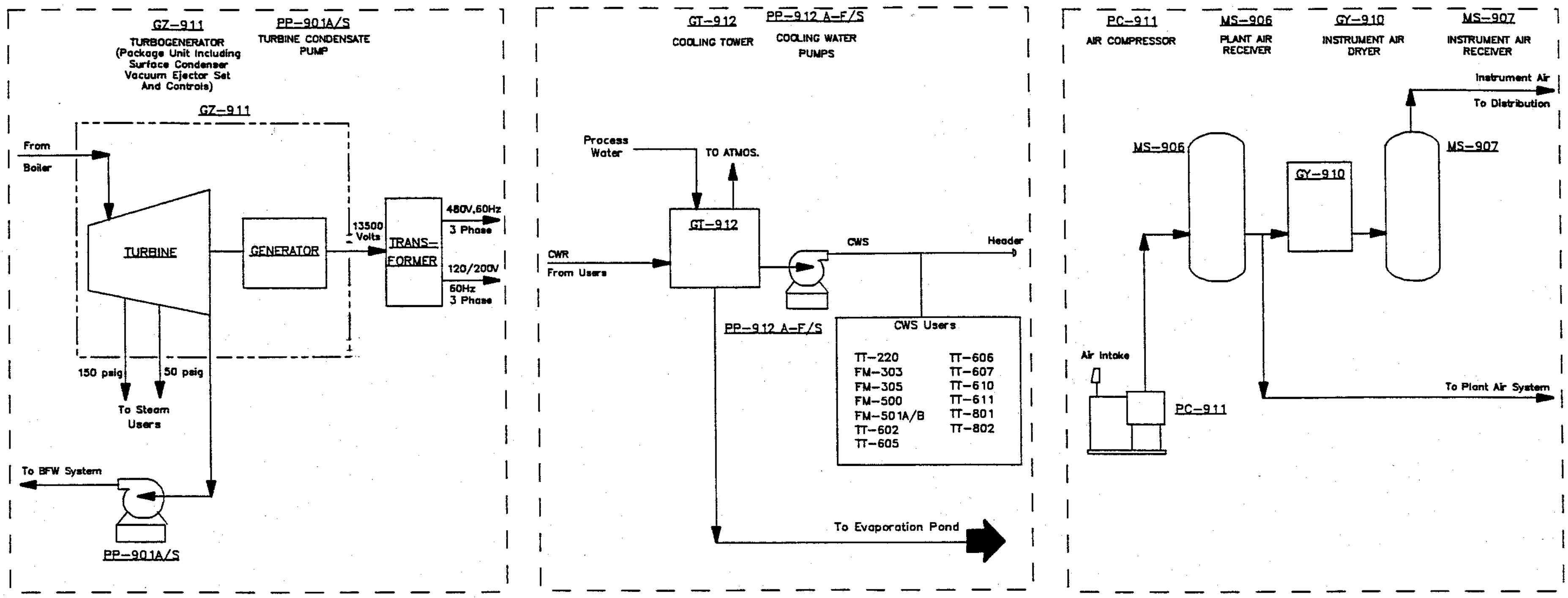

## Material Balance

| COMPONENTS/STREAM NO. (lb/hr) → | 822 | 921 | 922 |
|---|---|---|---|
| Water | 439,665 | 1,489,916 | 1,929,580 |
| Total | 439,665 | 1,489,916 | 1,929,580 |

◇ Stream No. ○ Stream Temperature(°C)

| | |
|---|---|
| Client DOE/Biofuels Program | SERI |
| Project No DOE/AFP/BESSF-10/89 | ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Draftsperson P. WALTER Date: | |
| Designer D.SCHELL N. HINMAN Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>UTILITIES<br>PROCESS WATER SYSTEM<br>AREA 900 |
| Approval N. HINMAN Date: | |
| Revision Code | SERI Drawing No.: BESSF-920 — Sheet 2 OF 6 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-19 — Date 9/27/90 |


T-930 CONDENSATE COLLECTION TANK
PP-910A/S CONDENSATE PUMP
GU-904A/S CONDENSATE POLISHER
GU-903A/B DEMINERALIZERS (PACKAGE UNIT)
MS-904 CONDENSATE SURGE DRUM
PP-909A/S DEAERATOR FEED PUMP
MS-903 HYDRAZINE DRUM
PP-907 HYDRAZINE TRANSFER PUMP
GU-907 HYDRAZINE ADDITION UNIT (PACKAGE UNIT)
GU-908 AMMONIA ADDITION UNIT (PACKAGE UNIT)
GU-909 PHOSPHATE ADDITION UNIT (PACKAGE UNIT)
GV-906 DEAERATOR
PP-908A/S BFW PUMP
MS-902 BLOWDOWN FLASH DRUM
PP-906A/S BLOWDOWN PUMP
GU-903A/B
Process Water
VENT TO ATMOS.
50 PSIG Steam
T-930
Cooled Condensate
LP Condensate
MP Condensate
GU-904A/S
MS-904
GV-906
MS-902
Blowdown From Boiler
To Boiler/HB-901A/B
PP-910A/S
PP-909A/S
PP-908A/S
To Aerobic Digestion/GV-808
PP-906A/S
To Boiler/HB-901A/B
Cond
Cond
Liquid Ammonia (From Rented Tank)
Cond
Phosphate (Dumped From Bags)
MS-903
GU-907
GU-908
GU-909
PP-907
Enclosed In Shed
Client
DOE/Biofuels Program
Project No
DOE/AFP/BESSF-10/89
Draftsperson P. WALTER Date:
Designer Date:
Approval N. HINMAN Date:
Revision Code
Misc. Data
SERI
ENGINEERING AND ANALYSIS SECTION
BIOTECHNOLOGY RESEARCH BRANCH
SOLAR FUELS RESEARCH DIVISION
Title CELLULOSIC BIOMASS-TO-ETHANOL
SIMULTANEOUS SACCHARIFICATION & FERMENTATION
UTILITIES
BFW SYSTEM
AREA 900
SERI Drawing No.: BESSF-930
Sheet 3 OF 6
Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-106-20
Date 6/21/90



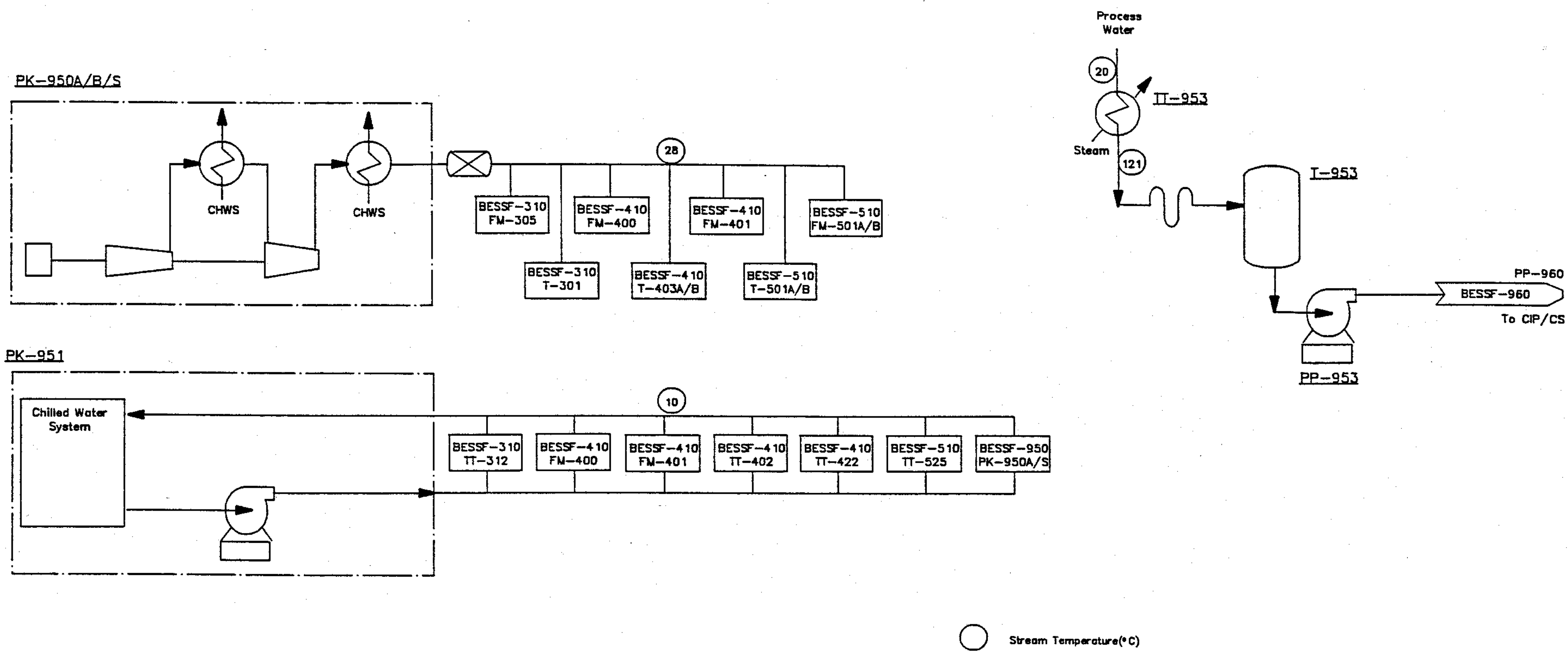

TURBOGENERATOR

COOLING WATER SYSTEM

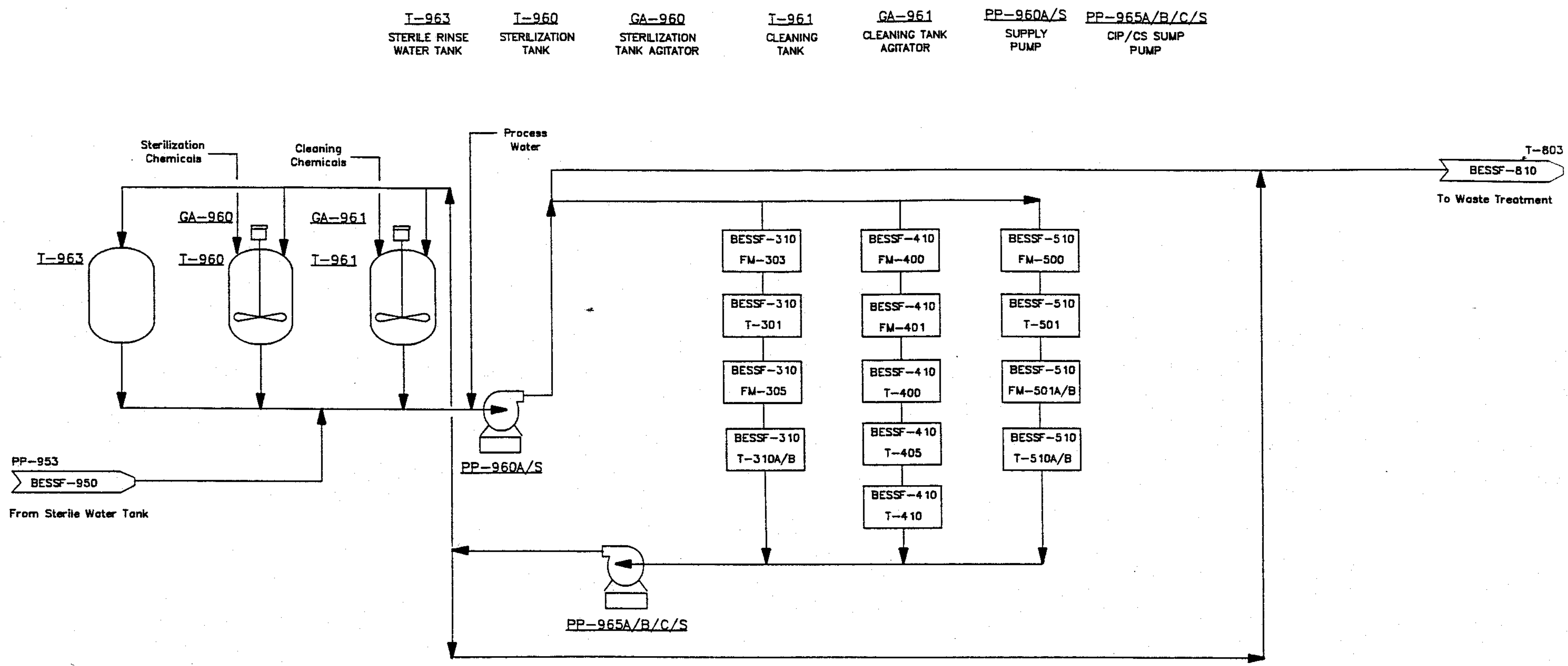
PLANT AND INSTRUMENT AIR

| | | |
|---|---|---|
| Client<br>DOE/Biofuels Program | SERI | |
| Project No<br>DOE/AFP/BESSF-10/89 | ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION | |
| Draftsperson Date:<br>P. WALTER | | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>UTILITIES<br>AREA 900 | |
| Approval Date:<br>N.HINMAN | | |
| Revision Code | SERI Drawing No.: BESSF-940 | Sheet 4 OF 6 |
| Misc. Data | Ref. Dwg. No. BADGER ENGINEERS,INC. E0461-106-21,22,23 | Date 9/26/90 |

| PK-950A/B/S | PK-951 | TT-953 | T-953 | PP-953 |
| --- | --- | --- | --- | --- |
| AIR COMPRESSOR PACKAGE | CHILLED WATER PACKAGE | WATER STERILIZER | STERILE WATER TANK | STERILE WATER PUMP |

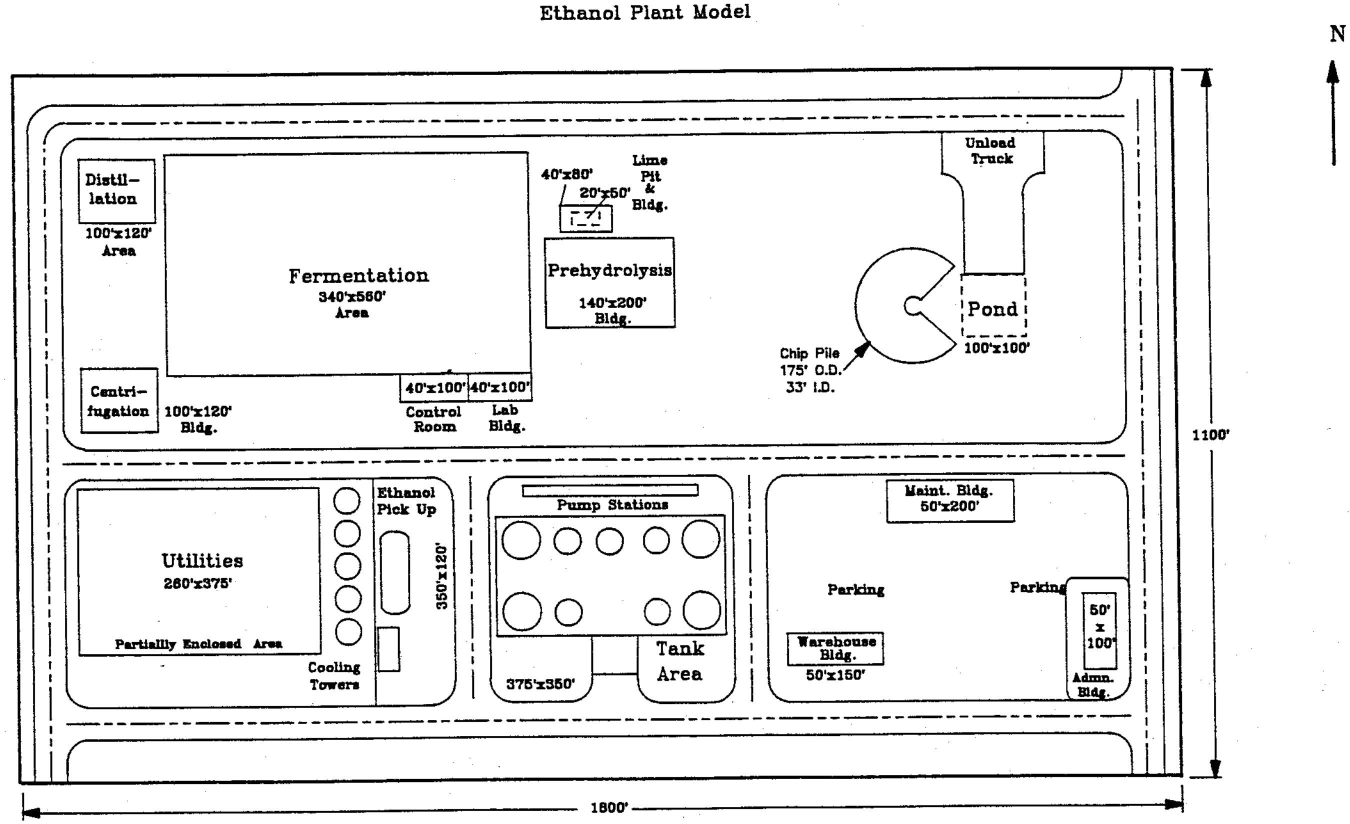



| | |
| --- | --- |
| Client<br>DOE/Biofuels Program | SERI |
| Project No<br>DOE/AFP/BESSF-10/89 | ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION |
| Approval Date:<br>N. HINMAN | AUXILIARY UTILITIES<br>AREA 900 |
| Revision Code | SERI Drawing No.: BESSF-950 — Sheet 5 of 6 |
| Misc. Data | Ref. Dwg. No. — Date 9/26/90 |

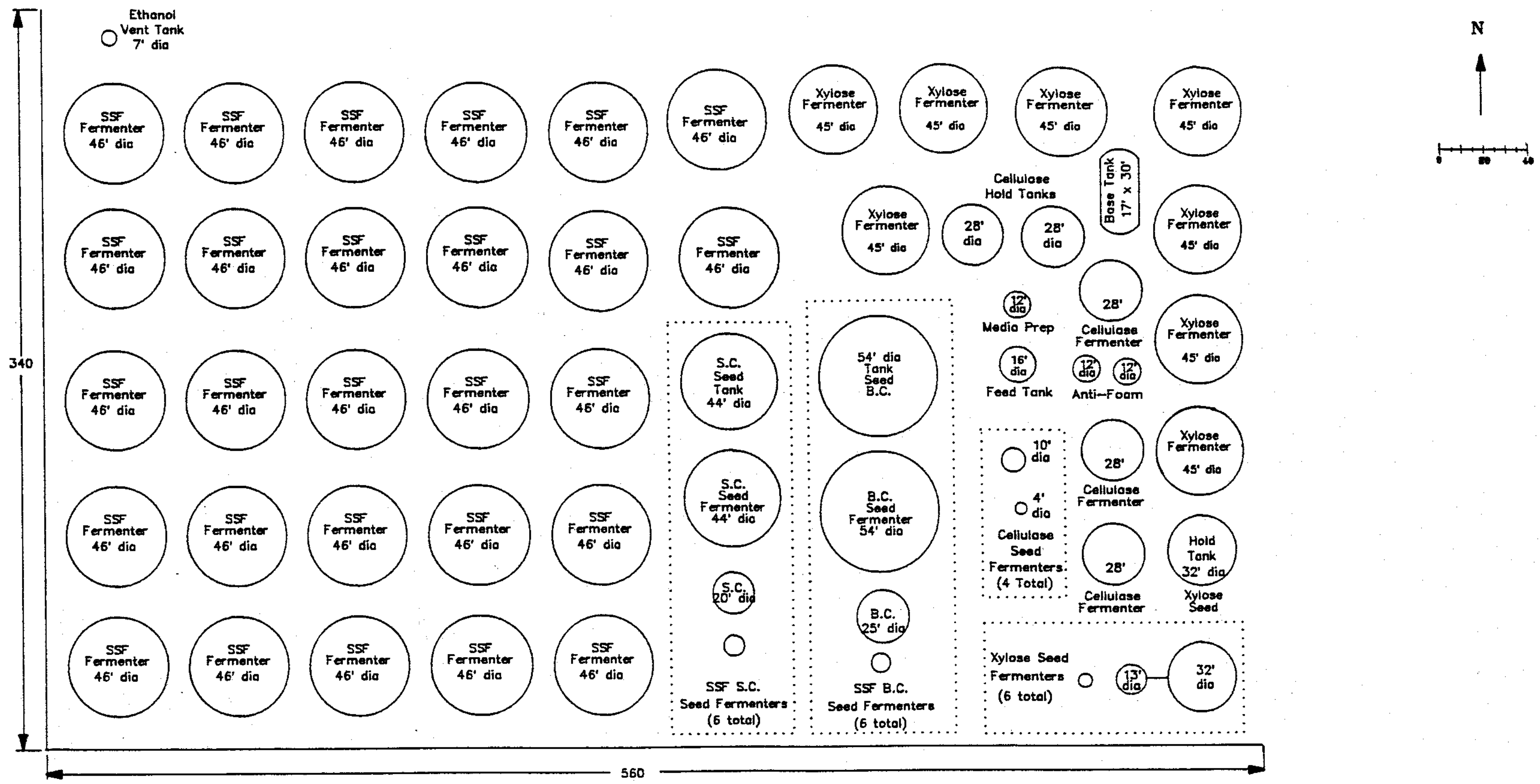



| | |
|---|---|
| Client<br>DOE/Biofuels Program | SERI<br>ENGINEERING AND ANALYSIS SECTION<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No<br>DOE/AFP/BESSF-10/89 | |
| Draftsperson Date:<br>P. WALTER | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION |
| Approval Date:<br>N. HINMAN | CLEAN IN PLACE & CHEMICAL STERILIZATION<br>AREA 900 |
| Revision Code | SERI Drawing No.: BESSF-960 — Sheet 6 of 6 |
| Misc. Data | Ref. Dwg. No. — Date 9/26/90 |

## Ethanol Plant Model




| | |
|---|---|
| Client DOE/Biofuels Program | SERI ENGINEERING AND ANALYSIS BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION |
| Project No DOE/AFP/BESSF-10/89 | |
| Draftsperson P. WALTER Date: | |
| Designer P. WALTER Date: | Title CELLULOSIC BIOMASS—TO—ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION Plot Plan |
| Approval N. HINMAN D. SCHELL Date: | |
| Revision Code | SERI Drawing No.: BESSF-1000 \| Sheet |
| Misc. Data 1" = 200' | Ref. Dwg. No. \| Date 8/28/90 |




| | |
|---|---|
| Client DOE/Biofuels Program | SERI ENGINEERING AND ANALYSIS<br>BIOTECHNOLOGY RESEARCH BRANCH<br>SOLAR FUELS RESEARCH DIVISION |
| Project No DOE/AFP/BESSF-10/89 | |
| Draftsperson P. WALTER Date: | |
| Designer Date: | Title CELLULOSIC BIOMASS-TO-ETHANOL<br>SIMULTANEOUS SACCHARIFICATION & FERMENTATION<br>FERMENTATION TANK LAYOUT |
| Approval N. HINMAN Date: | |
| Revision Code | SERI Drawing No.: BESSF-1001 — Sheet 1 OF 1 |
| Misc. Data | Ref. Dwg. No. — Date 8/27/90 |

Biomasss To Ethanol
Simultaneous Saccharification & Fermentation
Project No. DOE/AFP/BESSF/02

EQUIPMENT LIST
VESSELS

| ITEM # | EQUIPMENT NAME | ACTUAL FRACTION FULL | DIA/HGHT | PRESS. psi | TEMP. F | H V | MAT'L | Capacity gal | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T-201 | Sulfuric Acid Storage | 0.770 | 14ft/11ft | Atm | 100 | V | fiberglass | 5800 | BESSF-210 | 1 | $6,500 | $2,500 | Hall et al. 1995 |
| T-202 | Blowdown Tank | 0.499 | | 50 | 250 | V | SS 304 | 18000 | BESSF-210 | 1 | $50,000 | $50,000 | Icarus 1985 |
| T-206 | Neutralization Reaction Tank | 0.781 | 22ft/22ft | 50 | 250 | V | SS 304 | 24000 | BESSF-220 | 1 | $60,000 | $60,000 | Icarus 1985 |
| FM-302A-H | Xylose Fermenter | 0.944 | 45ft/60ft | Atm | 150 | V | CS | 746000 | BESSF-310 | 8 | $190,000 | $1,520,000 | Icarus 1985 |
| FM-305 | Xylose Seed Fermenter | 0.936 | 32ft/50ft | 50 | 150 | V | CS | 300000 | BESSF-310 | 1 | $88,000 | $68,000 | Icarus 1985 |
| FM-306 | Xylose Seed Fermenter | 0.941 | 14ft/25ft | 50 | 150 | V | CS | 29900 | BESSF-310 | 1 | $26,000 | $26,000 | Icarus 1985 |
| FM-307 | Xylose Seed Fermenter | 0.946 | 6.5ft/12ft | 50 | 150 | V | CS | 2990 | BESSF-310 | 1 | $8,000 | $8,000 | Icarus 1985 |
| FM-308 | Xylose Seed Fermenter | 0.898 | 3ft/5ft | 50 | 150 | V | CS | 310 | BESSF-310 | 1 | $3,400 | $3,400 | Ulrich 1984 |
| FM-309 | Xylose Seed Fermenter | 0.852 | 1.5ft/2.5ft | 50 | 150 | V | CS | 33 | BESSF-310 | 1 | $1,500 | $1,500 | Ulrich 1984 |
| FM-310 | Xylose Seed Fermenter | [illegible] | .7ft/1.5ft | 50 | 150 | V | CS | 4 | BESSF-310 | 1 | $700 | $700 | Ulrich 1984 |
| T-301 | Seed Hold Tank | [illegible] | 32ft/50ft | 10 | 150 | V | CS | 300000 | BESSF-310 | 1 | $85,000 | $85,000 | Icarus 1985 |
| T-321 | Base Tank | 0.869 | 17ft/30ft | 250 | 100 | H | CS | 28000 | BESSF-310 | 1 | $26,000 | $26,000 | Icarus 1985 |
| FM-400A/B/C | Cellulase Fermenter | 0.791 | 46ft/60ft | 15 | 150 | H | CS | 214000 | BESSF-410 | 3 | $65,000 | $195,000 | Icarus 1985 |
| FM-401A/B | Cellulase Seed Fermenter | 0.759 | 15ft/30ft | 50 | 250 | V | CS | 11000 | BESSF-410 | 2 | $20,000 | $40,000 | Icarus 1985 |
| FM-402A/B | Cellulase Seed Fermenter | 0.736 | 6ft/10ft | 50 | 250 | V | CS | 550 | BESSF-410 | 2 | $3,300 | $6,600 | Icarus 1985 |
| FM-403A/B | Cellulase Seed Fermenter | 0.801 | 2ft/4ft | 50 | 250 | V | CS | 26 | BESSF-410 | 2 | $1,100 | $2,200 | Icarus 1985 |
| FM-404A/B | Cellulase Seed Fermenter | 0.721 | .7/2ft | 50 | 250 | V | CS | 1.4 | BESSF-410 | 2 | $600 | $1,200 | Icarus 1985 |
| T-400 | Media Prep Tank | 0.773 | 12ft/18ft | 50 | 250 | V | CS | 9400 | BESSF-410 | 1 | $15,000 | $15,000 | Icarus 1985 |
| T-403A/B | Antifoam Tank | 0.828 | 10ft/10ft | 50 | 250 | V | CS | 840 | BESSF-410 | 2 | $4,000 | $6,000 | Ulrich 1984 |
| T-405 | Sterile Feed Tank | 0.788 | 16ft/30ft | 10 | 250 | V | CS | 183000 | BESSF-410 | 1 | $60,000 | $60,000 | Icarus 1985 |
| T-410A/B | Cellulase Hold Tank | | 46ft/60ft | Atm | 150 | V | CS | 214000 | BESSF-410 | 1 | $67,000 | $67,000 | Icarus 1985 |
| FM-500A-AA | SSF Fermenter | 0.948 | 46ft/60ft | Atm | 150 | V | CS | 778000 | BESSF-510 | 27 | $196,000 | $5,292,000 | Icarus 1985 |
| FM-501A | SSF Seed Fermenters (S.c.) | 0.946 | 43ft/60ft | 10 | 150 | V | CS | 451000 | BESSF-510 | 1 | $122,000 | $122,000 | Icarus 1985 |
| FM-502A | SSF Seed Fermenters (S.c.) | 0.946 | 19ft/30ft | 10 | 150 | V | CS | 45100 | BESSF-510 | 1 | $30,000 | $30,000 | Icarus 1985 |
| FM-503A | SSF Seed Fermenters (S.c.) | 0.927 | 8ft/16ft | 10 | 150 | V | CS | 4600 | BESSF-510 | 1 | $10,000 | $10,000 | Icarus 1985 |
| FM-504A | SSF Seed Fermenters (S.c.) | 0.848 | 4ft/7ft | 10 | 150 | V | CS | 500 | BESSF-510 | 1 | $3,400 | $3,400 | Ulrich 1984 |
| FM-505A | SSF Seed Fermenters (S.c.) | 0.806 | 2ft/3ft | 10 | 150 | V | CS | 70 | BESSF-510 | 1 | $1,500 | $1,500 | Ulrich 1984 |
| FM-506A | SSF Seed Fermenters (S.c.) | [illegible] | 1ft/1ft | 10 | 150 | V | CS | 4 | BESSF-510 | 1 | $500 | $500 | Ulrich 1984 |
| FM-501B | SSF Seed Fermenters (B.c.) | 0.931 | 52ft/60ft | 10 | 150 | V | CS | 953000 | BESSF-510 | 1 | $269,100 | $269,100 | Icarus 1985 |
| FM-502B | SSF Seed Fermenters (B.c.) | [illegible] | 20ft/38ft | 10 | 150 | V | CS | 89300 | BESSF-510 | 1 | $36,000 | $36,000 | Icarus 1985 |
| FM-503B | SSF Seed Fermenters (B.c.) | [illegible] | 10ft/15ft | 10 | 150 | V | CS | 9400 | BESSF-510 | 1 | $14,000 | $14,000 | Icarus 1985 |
| FM-504B | SSF Seed Fermenters (B.c.) | [illegible] | 5ft/7ft | 10 | 150 | V | CS | 1000 | BESSF-510 | 1 | $5,000 | $5,000 | Icarus 1985 |
| FM-505B | SSF Seed Fermenters (B.c.) | 0.757 | 2ft/4ft | 10 | 150 | V | CS | 94 | BESSF-510 | 1 | $1,600 | $1,600 | Ulrich 1984 |
| FM-506B | SSF Seed Fermenters (B.c.) | 0.507 | 1ft/2ft | 10 | 150 | V | CS | 12 | BESSF-510 | 1 | $900 | $900 | Ulrich 1984 |
| T-501A | Seed Hold Tank (S.c.) | 0.946 | 43ft/50ft | 10 | 150 | V | CS | 451000 | BESSF-510 | 1 | $122,000 | $122,000 | Icarus 1985 |
| T-501B | Seed Hold Tank (B.c.) | 0.931 | 52ft/60ft | 10 | 150 | V | CS | 764000 | BESSF-510 | 1 | $190,000 | $190,000 | Icarus 1985 |
| T-507 | Ethanol Vent Storage Tank | 0.781 | 7ft/20ft | 1.5 | 110 | V | CS | 4000 | BESSF-510 | 1 | $10,000 | $10,000 | Icarus 1985 |
| T-601 | Degasser Drum | 0.452 | 7ft/20ft | 15 | 300 | V | CS | 8200 | BESSF-610 | 1 | $14,000 | $14,000 | Icarus 1985 |
| T-602 | Beer Column Reflux Drum | 0.555 | 4ft/12ft | 50 | 250 | H | CS | 2500 | BESSF-610 | 1 | $8,000 | $8,000 | Icarus 1985 |
| T-603 | Fusel Oil Decanter | | 4ft/14ft | 25 | 150 | H | CS | 1300 | BESSF-620 | 1 | $6,000 | $6,000 | Icarus 1985 |
| T-605 | Rectification Column Rflux Drm | 0.592 | 7ft/20ft | 50 | 250 | H | CS | 2500 | BESSF-620 | 1 | $8,000 | $8,000 | Icarus 1985 |
| T-630 | Recycled Water Tank | 0.727 | | 50 | 250 | V | CS | 6800 | BESSF-630 | 1 | $15,000 | $15,000 | Icarus 1985 |
| T-701A/B | Ethanol Product Tank | 0.817 | 66ft/40ft | Atm | 150 | V | CS | 1023000 | BESSF-710 | 2 | $287,100 | $574,200 | Icarus 1985, Ulrich, 1982 |
| T-703 | Sulfuric Acid Storage Tank | 0.866 | 20ft/20ft | Atm | 150 | V | CS | 47000 | BESSF-710 | 1 | $28,000 | $28,000 | Icarus 1985 |
| T-704 | Fire Water Tank | | 52ft/32ft | Atm | 150 | V | CS | 508000 | BESSF-710 | 1 | $138,000 | $138,000 | Icarus 1985 |
| T-705A/B | NH3 Storage Tank | 0.840 | 16ft/40ft | 250 | 150 | H | CS | 59000 | BESSF-710 | 2 | $85,000 | $170,000 | Chemcost |
| T-70[illegible] | Antifoam Storage Tank | 0.731 | 5ft/7ft | Atm | 150 | V | CS | 2500 | BESSF-710 | 1 | $8,000 | $8,000 | Icarus 1985 |
| T-708 | Diesel Fuel Tank | | 15ft/16ft | Atm | 150 | V | CS | 21000 | BESSF-710 | 1 | $48,000 | $46,000 | Icarus 1985 |
| T-710 | Gasoline Storage Tank | | 30ft/32ft | Atm | 150 | V | CS | 169000 | BESSF-710 | 1 | $100,000 | $100,000 | Icarus 1985, Peters 1980 |
| T-720 | Corn Steep Liquor Tank | 0.935 | 20ft/15ft | Atm | 150 | V | CS | 36000 | BESSF-710 | 1 | $26,000 | $26,000 | Icarus 1985 |
| MS-805 | Reactor Surge Drum | | 15ft/20ft | 15 | 650 | V | CS | 16500 | BESSF-810 | 1 | $32,000 | $32,000 | Icarus 1985 |
| MS-806 | Offgas K.O. Suction Pot | | 8ft/11ft | 15 | 650 | V | CS | 2800 | BESSF-810 | 1 | $8,000 | $8,000 | Icarus 1985 |
| MS-809 | Offgas Knock Out Drum | | 8ft/11ft | 15 | 650 | V | CS | 2800 | BESSF-810 | 1 | $8,000 | $8,000 | Icarus 1985 |
| MS-810 | LP Vent Knock Out Drum | | 3ft/9ft | 15 | 150 | V | CS | 470 | BESSF-830 | 1 | $4,500 | $4,500 | Ulrich 1984 |
| T-803 | Equalization Tank | | 82ft/40ft | Atm | 70 | V | CS | 962000 | BESSF-810 | 1 | $236,000 | $236,000 | Icarus 1985 |
| T-804 | Anaerobic Reactor | | 30ft/50ft | 15 | 650 | V | CS | 264000 | BESSF-810 | 1 | $79,000 | $79,000 | Icarus 1985 |
| T-807 | Biotreater | | 74ft/30ft | Atm | 70 | V | Epoxy/CS | 552000 | BESSF-820 | 1 | $161,000 | $161,000 | Icarus 1985 |
| MS-902 | Blowdown Flash Drum | | 3ft/10ft | 50 | 250 | H | CS | 520 | BESSF-930 | 1 | $4,500 | $4,500 | Ulrich 1984 |
| MS-903 | Hydrazine Drum | | 3ft/5ft | 10 | 250 | V | SS 316 | 260 | BESSF-930 | 1 | $8,200 | $8,200 | Icarus 1985 |
| MS-904 | Condensate Surge Drum | | 10ft/30ft | 15 | 250 | H | CS | 11400 | BESSF-930 | 1 | $24,000 | $24,000 | Icarus 1985 |
| MS-906 | Plant Air Receiver | | 5ft/15ft | 150 | 150 | V | CS | 2300 | BESSF-940 | 1 | $27,500 | $27,500 | Icarus 1985 |
| MS-907 | Instrument Air Receiver | | 5ft/15ft | 150 | 150 | V | CS | 2300 | BESSF-940 | 1 | $27,500 | $27,500 | Icarus 1985 |
| T-901 | Process Water Tank | | 50ft/40ft | Atm | 150 | V | CS | 587000 | BESSF-920 | 1 | $154,000 | $154,000 | Icarus 1985 |
| T-905 | Backwash Transfer Tank | | 22ft/4ft | Atm | 70 | V | CS | 11000 | BESSF-920 | 1 | $20,000 | $20,000 | Icarus 1985 |
| T-930 | Condensate Collection Tank | | 6ft/10ft | 250 | 500 | V | CS/rubber | 1600 | BESSF-930 | 1 | $53,200 | $53,200 | Icarus 1985 |
| T-952 | Sterile Water Tank | | 10ft/17ft | Atm | 200 | V | CS | 10000 | BESSF-950 | 1 | $14,000 | $14,000 | Icarus 1985 |
| T-950 | Sterilization Tank | | 10ft/17ft | Atm | 200 | V | SS 304 | 10000 | BESSF-960 | 1 | $32,000 | $32,000 | Icarus 1985 |
| T-951 | Cleaning Tank | | 10ft/17ft | Atm | 200 | V | CS | 10000 | BESSF-950 | 1 | $14,000 | $14,000 | Icarus 1985 |
| T-953 | Sterile Rinse Water Tank | | 10ft/17ft | Atm | 200 | V | CS | 10000 | BESSF-950 | 1 | $14,000 | $14,000 | Icarus 1985 |

Total Equipment Cost $10,448,700

# Appendix F

## Equipment List

This section contains the equipment list for all equipment shown on the process flow diagrams. The list gives the equipment number, equipment name, specifications, cost, and source of the cost information.

Biomass To Ethanol
Simultaneous Saccharification & Fermentation
Project No. DOE/AFP/BESSF/02

E Q U I P M E N T   L I S T
PUMPS

| ITEM # | PUMP NAME | LB/HR | LB/CUFT | GPM | CAPICITY GPM | DELTA P psi | HYDRAUL. HP | PUMP EFFIC. | BRAKE HP | MOTOR HP/UNIT | MAT'L | TYPE | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| PP-101 | Flume Pump | 1000000 | 62.4 | 1998.0 | 2000 | 5 | 5.83 | 0.75 | 7.78 | 10 | Cast Steel | Centrifugal | BESSF-110 | 1 | $4,300 | $4,300 | Icarus 1985 |
| PP-201A/S | Sulfuric Acid Pump | 2878 | 112.4 | 3.2 | 4 | 32 | 0.07 | 0.4 | 0.19 | 1 | SS | Reciprocating | BESSF-210 | 2 | $3,200 | $6,400 | Icarus 1987 |
| PP-202A/S | Hydrolyzate Pump | 942705 | 62.4 | 1883.5 | 2000 | 20 | 23.34 | 0.75 | 31.12 | 50 | SS | Pos. Disp. | BESSF-210 | 2 | $67,000 | $134,000 | Icarus 1985 |
| PP-203A/S | Neutralized Hydrolyzate | 944833 | 62.4 | 1887.8 | 2000 | 50 | 58.34 | 0.75 | 77.79 | 100 | CS | Pos. Disp. | BESSF-210 | 2 | $63,000 | $126,000 | Icarus 1985 |
| PP-401A/S | Feed Pump | 20125 | 62.4 | 40.2 | 525 | 20 | 6.13 | 0.7 | 8.75 | 15 | CS | Pos. Disp. | BESSF-410 | 2 | $13,000 | $26,000 | Icarus 1985 |
| PP-403A/S | Fermenter Recycle Pump | | | | 750 | 50 | 21.88 | 0.7 | 31.26 | 40 | CS | Centrifugal | BESSF-410 | 2 | $6,600 | $13,200 | Icarus 1985 |
| PP-411A/S | Prep Tank Transfer Pump | 590 | 62.4 | 1.2 | 18 | 25 | 0.26 | 0.4 | 0.66 | 1 | CS | Centrifugal | BESSF-410 | 2 | $2,100 | $4,200 | Icarus 1985 |
| PP-412A/S | Cellulase Feed Pump | 29438 | 62.4 | 58.8 | 65 | 50 | 1.90 | 0.45 | 4.21 | 7.5 | CS | Centrifugal | BESSF-410 | 2 | $3,300 | $6,600 | Icarus 1985 |
| PP-505A/S | Beer Transfer Pump | 957399 | 62.4 | 1912.9 | 2000 | 30 | 35.01 | 0.75 | 46.67 | 60 | CS | Centrifugal | BESSF-510 | 2 | $7,500 | $15,000 | Icarus 1985 |
| PP-509A/S | Ethanol Vent Pump | 5259 | 49 | 13.4 | 15 | 60 | 0.53 | 0.4 | 1.31 | 2 | CS | Centrifugal | BESSF-510 | 2 | $2,600 | $5,200 | Icarus 1985 |
| PP-601A/S | Beer Column Bottoms Pump | 916659 | 62.4 | 1831.5 | 2000 | 60 | 70.01 | 0.75 | 93.35 | 125 | CS | Centrifugal | BESSF-610 | 2 | $9,700 | $19,400 | Icarus 1985 |
| PP-603A/S | Beer Column Reflux Pump | 141029 | 62.4 | 281.8 | 300 | 60 | 10.50 | 0.65 | 16.16 | 20 | CS | Centrifugal | BESSF-610 | 2 | $5,000 | $10,000 | Icarus 1985 |
| PP-604A/S | Wash Return Pump | | | | 25 | 60 | 0.88 | 0.4 | 2.19 | 3 | CS | Centrifugal | BESSF-620 | 2 | $3,000 | $6,000 | Icarus 1985 |
| PP-605A/S | Fusel Oil Pump | 115 | 62.4 | 0.2 | 0.25 | 40 | 0.01 | 0.3 | 0.02 | 1 | CS | Centrifugal | BESSF-620 | 2 | $2,100 | $4,200 | Icarus 1985 |
| PP-607A/S | Rectification Column Btms | 60615 | 62.4 | 121.1 | 130 | 40 | 3.03 | 0.45 | 6.74 | 10 | CS | Centrifugal | BESSF-620 | 2 | $4,300 | $8,600 | Icarus 1985 |
| PP-608A/S | Rectification Column Rflx | 118014 | 49 | 300.3 | 320 | 60 | 11.20 | 0.65 | 17.23 | 25 | CS | Centrifugal | BESSF-620 | 2 | $5,600 | $11,200 | Icarus 1985 |
| PP-631A/S | Recycled Water Pump | 802129 | 62.4 | 1602.7 | 1700 | 30 | 29.75 | 0.75 | 39.67 | 50 | CS | Centrifugal | BESSF-630 | 2 | $7,000 | $14,000 | Icarus 1985 |
| PP-632A/S | Sump Pump | | | | 100 | 20 | 1.17 | 0.45 | 2.59 | 5 | CS | Centrifugal | BESSF-630 | 2 | $3,300 | $6,600 | Icarus 1985 |
| PP-701A/B/S | Ethanol Export Pump | | | | 525 | 30 | 9.19 | 0.7 | 13.13 | 20 | CS | Centrifugal | BESSF-710 | 3 | $5,000 | $15,000 | Icarus 1985 |
| PP-703A/S | Sulfuric Acid Transfer | | | | 100 | 40 | 2.33 | 0.45 | 5.19 | 7.5 | SS | Reciprocating | BESSF-710 | 2 | $17,700 | $35,400 | Chemcost |
| PP-704A/S | Fire Water Pump | | | | 600 | 100 | 35.01 | 0.7 | 50.01 | 60 | CS | Centrifugal | BESSF-710 | 2 | $7,500 | $15,000 | Icarus 1985 |
| PP-706A/S | NH3 Transfer Pump | | | | 200 | 40 | 4.67 | 0.6 | 7.78 | 10 | CS | Centrifugal | BESSF-710 | 2 | $4,300 | $8,600 | Icarus 1985 |
| PP-707A/S | Antifoam Transfer Pump | | | | 10 | 40 | 0.23 | 0.4 | 0.58 | 1 | CS | Centrifugal | BESSF-710 | 2 | $2,100 | $4,200 | Icarus 1985 |
| PP-708A/S | Diesel Fuel Pump | | | | 25 | 50 | 0.73 | 0.4 | 1.82 | 3 | CS | Centrifugal | BESSF-710 | 2 | $2,900 | $5,800 | Icarus 1985 |
| PP-710A/S | Gasoline Blending Pump | | | | 7.5 | 30 | 0.13 | 0.4 | 0.33 | 1 | CS | Centrifugal | BESSF-710 | 2 | $2,100 | $4,200 | Icarus 1985 |
| PP-720A/S | Corn Steep Liquor Transfr | | | | 60 | 60 | 2.10 | 0.45 | 4.67 | 7.5 | SS | Centrifugal | BESSF-710 | 2 | $3,500 | $7,000 | Icarus 1985 |
| PP-808A/S | Reactor Feed Pump | | | | 1100 | 50 | 32.09 | 0.75 | 42.78 | 60 | CS | Centrifugal | BESSF-810 | 2 | $7,500 | $15,000 | Icarus 1985 |
| PP-809 | Reactor Recycle Pump | | | | 1100 | 50 | 32.09 | 0.75 | 42.78 | 60 | CS | Centrifugal | BESSF-810 | 1 | $7,500 | $7,500 | Icarus 1985 |
| PP-813A/S | Sludge Pump | | | | 265 | 20 | 3.09 | 0.65 | 4.76 | 7.5 | CS | Pos. Displacmnt | BESSF-820 | 2 | $9,200 | $18,400 | Icarus 1985 |
| PP-816A/S | Final Effluent Pump | | | | 910 | 40 | 21.24 | 0.75 | 28.32 | 40 | CS | Centrifugal | BESSF-820 | 2 | $6,500 | $13,000 | Icarus 1985 |
| PP-901A/S | Turbine Condensate Pump | | | | 200 | 40 | 4.67 | 0.6 | 7.78 | 10 | CS | Centrifugal | BESSF-940 | 2 | $4,300 | $8,600 | Icarus 1985 |
| PP-902A/S | Process Water Transfer | | | | 2650 | 30 | 46.38 | 0.75 | 61.84 | 75 | CS | Centrifugal | BESSF-920 | 2 | $8,000 | $16,000 | Icarus 1985 |
| PP-903A/S | Process Water Circulating | | | | 2650 | 60 | 92.77 | 0.75 | 123.69 | 150 | CS | Centrifugal | BESSF-920 | 2 | $10,200 | $20,400 | Icarus 1985 |
| PP-904A/B | Backwash Feed Pump | | | | 6000 | 20 | 70.01 | 0.8 | 87.51 | 125 | CS | Centrifugal | BESSF-920 | 2 | $9,700 | $19,400 | Icarus 1985 |
| PP-905A/B - | Backwash Transfer Pump | | | | 50 | 20 | 0.58 | 0.4 | 1.46 | 2 | CS | Centrifugal | BESSF-920 | 2 | $2,600 | $5,200 | Icarus 1985 |
| PP-906A/S | Blowdown Pump | | | | 50 | 25 | 0.73 | 0.4 | 1.82 | 3 | CS | Centrifugal | BESSF-930 | 2 | $2,900 | $5,800 | Icarus 1985 |
| PP-907 | Hydrazine Transfer Pump | | | | 5 | 20 | 0.06 | 0.35 | 0.17 | 1 | SS | Centrifugal | BESSF-930 | 1 | $2,200 | $2,200 | Icarus 1985 |
| PP-908A/S | Boiler Feed Water Pump | | | | 1100 | 1250 | 802.22 | 0.75 | 1069.62 | 1200 | SS | Cntrfgl,staged | BESSF-930 | 2 | $95,000 | $190,200 | Icarus 1985 |
| PP-909A/S | Dearator Feed Pump | | | | 1100 | 20 | 12.84 | 0.75 | 17.11 | 25 | CS | Centrifugal | BESSF-930 | 2 | $5,600 | $11,200 | Icarus 1985 |
| PP-910A/S | Condensate Pump | | | | 1100 | 30 | 19.25 | 0.75 | 25.67 | 40 | CS | Centrifugal | BESSF-940 | 2 | $6,500 | $13,000 | Icarus 1985 |
| PP-912A-F/S | Cooling Water Pumps | | | | 9000 | 60 | 315.05 | 0.8 | 393.82 | 500 | CS | Centrifugal | BESSF-940 | 7 | $15,600 | $109,200 | Icarus 1985 |
| PP-913A/S | Well Water Pumps | | | | 1800 | 25 | 26.25 | 0.75 | 35.01 | 50 | CS | Centrifugal | BESSF-920 | 2 | $7,000 | $14,000 | Icarus 1985 |
| PP-953 | Sterile Water Pump | | | | 10 | 30 | 0.18 | 0.4 | 0.44 | 1 | CS | Centrifugal | BESSF-950 | 1 | $2,100 | $2,100 | Icarus 1985 |
| PP-960A/S | Supply Pump | | | | 20 | 50 | 0.58 | 0.4 | 1.46 | 2 | SS | Centrifugal | BESSF-960 | 2 | $3,000 | $6,000 | Icarus 1985 |
| PP-965A/B/C/S | CIP/CS Sump Pump | | | | 20 | 30 | 0.35 | 0.4 | 0.88 | 2 | SS | Centrifugal | BESSF-960 | 4 | $3,000 | $12,000 | Icarus 1985 |

PUMPS
Total Cost: $1,005,100
Total HP Req'd: 5190

Biomass To Ethanol
Simultaneous Saccharification & Fermentation
Project No. DOE/AFP/BESSF/02

E Q U I P M E N T   L I S T
SOLIDS HANDLING

| ITEM # | EQUIPMENT NAME | TYPE | DUTY/DESCRIPTION | HP/ UNIT | MAT'L | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|
| GS-103 | Magnetic Chip Cleaner | | remove down to .5" nuts | 7 | CS | BESSF-110 | 1 | $10,300 | $10,300 | Icarus 1987 |
| GM-101A/B/S | Front End Loaders | Diesel | | | | BESSF-110 | 3 | $156,000 | $468,000 | Icarus 1987 |
| GS-101 | Radial Stacking Conveyer | Paddle | 1500 t/h wet wght, 120ft x 30in | 15 | CS | BESSF-110 | 1 | $124,900 | $124,900 | Morbark 1983 |
| GS-102 | Belt Conveyer | Belt | 240ft x 6.5ft wide, 200 t/h | 10 | CS | BESSF-110 | 1 | $191,700 | $191,700 | Ulrich 1984 |
| GS-104 | Milled Chip Belt Conveyor | Belt | 50ft X 6.5ft wide, 200 t/h | 5 | CS | BESSF-110 | 1 | $40,000 | $40,000 | Ulrich 1984 |
| GY-101A/B/C/D | Wood Chip Unloader with Scale | 23-ton/load | 10 vans/hr/loader | 35 | CS | BESSF-110 | 4 | $39,400 | $157,600 | Morbark 1983 |
| GG-101A/B/C/D | Disk Refiner | | | 2500 | CS | BESSF-110 | 4 | $379,000 | $1,516,000 | Sprout-Bauer Quote, 1990 |
| GS-202 | Screw Feeder | Auger | 19500 cft/h | 100 | CS | BESSF-210 | 2 | $280,000 | $560,000 | Black Clawson Quote 1990 |
| GS-223 | Lime Unloading Conveyer | Bucket | 120 ft high, 100t/h | 50 | CS | BESSF-220 | 1 | $18,000 | $18,000 | Peters et al.80, Ulrich 84 |
| GS-225 | Lime Solids Feeder | Rotary Vlv | 1.5 t/h | 1 | CS | BESSF-220 | 1 | $5,000 | $5,000 | |
| MB-220 | Lime Storage Bin | | 3500 cft | | CS | BESSF-220 | 1 | $17,000 | $17,000 | Icarus 1985 |
| MF-224 | Lime Unloading Pit | | 605 cft, 20ft X 50ft X 3ft | | Concrete | BESSF-220 | 1 | $5,000 | $5,000 | Means 1987 |
| GS-611A/B | Sludge Screws | Screw, 18in | 150ft long | 10 | CS | BESSF-630 | 2 | $20,500 | $41,000 | Ulrich 1984 |
| GS-801 | Sludge Screws | Screw, 9in | 100ft long | 1.5 | CS | BESSF-820 | 1 | $8,000 | $8,000 | Ulrich 1984 |

SOLIDS HANDLING EQUIPMENT
Total Cost: $3,162,500
Total HP Req'd: 10449.5

Biomass To Ethanol
Simultaneous Saccharification & Fermentation
EAS Project No. DOE/AFP/BESSF/02

EQUIPMENT LIST
HEAT EXCHANGER

| ITEM # | EQUIPMENT NAME | AREA SQ. FT. | TUBE MAT'L/DP psig /TEMP deg F | SHELL MAT'L/DP psig /TEMP deg F | TYPE | REMARKS | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|---|
| TT-220 | Feed Cooler | 11362 | CS/100 | CS/100 | Fixed Tube | Single pass | BESSF-220 | 1 | $22,600 | $22,600 | Ulrich 1984 |
| TT-312 | Exhaust Condenser | 1265 | CS/100 | CS/100 | Vent Condenser | | BESSF-310 | 1 | $14,100 | $14,100 | Hall et al. 1982 |
| TT-402 | Water Cooler | 80 | CS/100 | CS/100 | Fixed Tube | Single pass | BESSF-410 | 1 | $3,000 | $3,000 | Hall et al. 1968 |
| TT-422 | Fermenter Exhaust Cndnsr | 605 | CS/100 | CS/100 | Vent Condenser | | BESSF-410 | 1 | $3,100 | $3,100 | Hall et al. 1982 |
| TT-525 | Exhaust Condenser | 1800 | CS/100 | CS/100 | Vent Condenser | | BESSF-510 | 2 | $17,600 | $35,200 | Hall et al. 1982 |
| TT-602 | Degasser Drum Condenser | 25 | CS/100 | CS/100 | Vent Condenser | | BESSF-610 | 1 | $3,400 | $3,400 | Hall et al. 1982 |
| TT-603 | Beer Column Reboiler | 1743 | CS/100 | CS/100 | Reboiler | | BESSF-610 | 1 | $58,800 | $58,800 | Peters 1968, ICARUS |
| TT-605 | Beer Column Condenser | 1400 | CS/100 | CS/100 | Fixed tube | | BESSF-610 | 2 | $15,600 | $31,200 | ICARUS 1985 |
| TT-606 | Beer Column Vent Condensr | 36 | CS/100 | CS/100 | Vent Condenser | | BESSF-610 | 1 | $3,400 | $3,400 | Hall et al. 1982 |
| TT-607 | Fusel Oil Cooler | 4 | CS/100 | CS/100 | Double pipe | | BESSF-620 | 1 | $400 | $400 | |
| TT-609 | Rectification Clmn Rboilr | 980 | CS/100 | CS/100 | Reboiler | | BESSF-620 | 1 | $20,800 | $20,800 | Icarus 1985 |
| TT-610 | Rectification Clmn Cndsr | 1785 | CS/100 | CS/100 | Fixed tube | | BESSF-620 | 1 | $20,800 | $20,800 | Hall et al. 1988 |
| TT-611 | Rctfctn Clmn Vent Cndsr | 43 | CS/100 | CS/100 | Vent Condenser | | BESSF-620 | 1 | $3,400 | $3,400 | Hall et al. 1982 |
| TT-613 | Feed Preheater | 750 | CS/100 | CS/100 | Fixed tube | | BESSF-610 | 1 | $9,600 | $9,600 | Hall et al. 1988, Pe |
| TT-615 | Feed Cross Exchanger | 1331 | CS/100 | CS/100 | Floating head | | BESSF-610 | 1 | $35,400 | $35,400 | Hall et al. 1988. Pe |
| TT-801 | Offgas Cooler | 500 | CS/100 | CS/100 | Fixed tube | | BESSF-810 | 1 | $8,000 | $8,000 | Hall et al. 1988 |
| TT-802 | Feed Cooler | 1600 | CS/100 | CS/100 | Fixed tube | 4 shell-8 tube pass | BESSF-810 | 5 | $16,600 | $83,000 | Hall et al. 1988 |
| TT-953 | Water Sterilizer | 1.25 | CS/100 | CS/100 | Double pipe | | BESSF-950 | 1 | $200 | $200 | |
| | Xylose Fer. Coils | 2257 | CS/100 | | Coils | 4.0 in OD, Sch 10 | BESSF-310 | 8 | $2,900.00 | $23,200 | Vendor Quote |
| | Cellulase Fer. Coils | 8627 | CS/100 | | Coils | 6.0 in OD, Sch 10 | BESSF-410 | 3 | $13,900 | $41,700 | Vendor Quote |
| | SSF Fer. Coils | 692 | CS/100 | | Coils | 2.5 in OD, Sch 10 | BESSF-510 | 27 | $1,000 | $27,000 | Vendor Quote |
| | | | | | | | | | Total Equipment Cost | $448,300 | |

Biomass To Ethanol
Simultaneous Saccharification & Fermentation
Project No. DOE/AFP/BESSF/02

EQUIPMENT LIST
MISCELLANEOUS

| ITEM # | EQUIPMENT NAME | DUTY/DESCRIPTION | HP/ Unit | MAT'L | REMARKS | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|
| GA-201 | Line Mixer | | | Hastalloy | | BESSF-210 | 1 | $2,100 | $2,100 | Icarus 1987 |
| GA-203 | Blowdown Tank Agitator | Single Impeller | 25 | SS 304 | | BESSF-210 | 1 | $18,300 | $18,300 | Chemcost |
| GA-213 | Neutralization Tank Agitator | Single Impeller | 50 | SS 304 | | BESSF-220 | 1 | $29,600 | $29,600 | Chemcost |
| GC-227 | Lime Dust Cyclone | 250 ASCFM, 15 lb/h solids | | CS | | BESSF-220 | 1 | $800 | $700 | Chemcost |
| GF-201 | Desiccant Air Filter | 0.5 cfm | | Silica | | BESSF-210 | 1 | $1,000 | $1,000 | |
| MR-201 | Impregnator with Rotary Valve | 2564 cu ft, 20 HP drive, 20 HP rtry vlve | 40 | C-20 Cb3 | Continuous pulp digestor | BESSF-210 | 2 | $1,830,000 | $3,660,000 | Black Clawson 1990 |
| MR-202 | Prehydrolysis Reactor | 2564 cu ft, 20 HP drive, 20 HP rtry vlve | 40 | C-20 Cb3 | Continuous pulp digestor | BESSF-210 | 2 | $1,830,000 | $3,660,000 | Black Clawson 1990 |
| GA-301 | Seed Hold Tank Agitator | Single Impeller/ 40 hp | 20 | CS | | BESSF-310 | 1 | $16,900 | $16,900 | Chemcost |
| GA-303A-H | Xylose Fermenter Agitator | Single Impeller | 75 | CS | | BESSF-310 | 8 | $28,900 | $231,200 | Chemcost |
| GA-305 | First Seed Vessel Agitator | Single Impeller | 150 | CS | | BESSF-310 | 1 | $55,200 | $55,200 | Chemcost |
| GA-306 | Second Seed Vessel Agitator | Single Impeller | 40 | CS | | BESSF-310 | 1 | $16,900 | $16,900 | Chemcost |
| GA-307 | Third Seed Vessel Agitator | Single Impeller | 5 | CS | | BESSF-310 | 1 | $6,000 | $6,000 | Chemcost |
| GA-308 | Fourth Seed Vessel Agitator | Single Impeller | 0.5 | CS | | BESSF-310 | 1 | $2,000 | $2,000 | Chemcost |
| GA-400 | Prep Tank Agitator | Single Impeller/ 10 hp | 0.6 | CS | | BESSF-410 | 1 | $7,600 | $7,600 | Chemcost |
| GA-401A/B/C | Fermenter Agitator | Single Impeller | 200 | CS | | BESSF-410 | 3 | $75,300 | $225,900 | Chemcost |
| GA-405 | Feed Tank Agitator | Single Impeller/ 300 hp | 150 | CS | | BESSF-410 | 1 | $127,000 | $127,000 | Chemcost |
| GA-410 | Hold Tank Agitator | Single Impeller/ 200 hp | 100 | CS | | BESSF-410 | 1 | $75,300 | $75,300 | Chemcost |
| GA-411 | First Seed Vessel Agitator | Single Impeller | 40 | CS | | BESSF-410 | 2 | $16,900 | $33,800 | Chemcost |
| GA-412 | Second Seed Vessel Agitator | Single Impeller | 1.5 | CS | | BESSF-410 | 2 | $3,000 | $6,000 | Chemcost |
| GA-413 | Third Seed Vesssel Agitator | Single Impeller | 0.25 | CS | | BESSF-410 | 2 | $1,500 | $3,000 | Chemcost |
| GA-500A-AA | SSF Fermenter Agitator | Single Impeller | 80 | CS | | BESSF-510 | 27 | $29,900 | $807,300 | Chemcost |
| GA-501A | Seed Hold Tank Agitator (S.c.) | Single Impeller/ 50 hp | 25 | CS | | BESSF-510 | 1 | $20,100 | $20,100 | Chemcost |
| GA-501B | Seed Hold Tank Agitator (B.c.) | Single Impeller/ 75 hp | 35 | CS | | BESSF-510 | 1 | $28,900 | $28,900 | Chemcost |
| GA-510A | First Seed Vessel Agitator (S.c.) | Single Impeller | 250 | CS | | BESSF-510 | 1 | $97,200 | $97,200 | Chemcost |
| GA-511A | Second Seed Vessel Agitator (S.c.) | Single Impeller | 50 | CS | | BESSF-510 | 1 | $20,100 | $20,100 | Chemcost |
| GA-512A | Third Seed Vessel Agitator (S.c.) | Single Impeller | 7.5 | CS | | BESSF-510 | 1 | $6,800 | $6,800 | Chemcost |
| GA-513A | Fourth Seed Vessel Agitator (S.c.) | Single Impeller | 0.75 | CS | | BESSF-510 | 1 | $2,500 | $2,500 | |
| GA-510B | First Seed Vessel Agitator (B.c.) | Single Impeller | 400 | CS | | BESSF-510 | 1 | $172,600 | $172,600 | Chemcost |
| GA-511B | Second Seed Vessel Agitator (B.c.) | Single Impeller | 75 | CS | | BESSF-510 | 1 | $28,900 | $28,900 | Chemcost |
| GA-512B | Third Seed Vessel Agitator (B.c.) | Single Impeller | 10 | CS | | BESSF-510 | 1 | $7,500 | $7,600 | Chemcost |
| GA-513B | Fourth Seed Vessel Agitator (B.c.) | Single Impeller | 1 | CS | | BESSF-510 | 1 | $2,700 | $2,720 | |
| GC-609A/B/C | Centrifuge | Solid Bowl | 200 | CS | Assumed 75% SS cost | BESSF-630 | 3 | $225,000 | $675,000 | Badger 1984 |
| GF-703 | Desiccant Air Filter | 1500 cfm | | Silica | | BESSF-710 | 1 | $38,800 | $38,800 | Icarus 1985 |
| GC-801 | Sludge Centrifuge | Solid Bowl | 40 | CS | Assumed 75% SS cost | BESSF-820 | 1 | $128,000 | $128,000 | Badger 1984 |
| GO-806 | Offgas Burner | | | CS | | BESSF-810 | 1 | $20,000 | $20,000 | |
| GV-807 | Biotreater Agitators | | | Polyethylene | | BESSF-820 | 1 | $10,000 | $10,000 | |
| GV-808 | Secondary Clarifier | Center feed, 100 ft diameter | 3 | CS | | BESSF-820 | 1 | $260,000 | $260,000 | Ulrich 1984 |
| PB-810 | Offgas Blower | 2630 cfm, 20 psig discharge | 60 | CS | | BESSF-810 | 1 | $74,300 | $74,300 | Peters and Timmernaus 1980 |
| PB-812A/S | Aeration Blowers | 1300 cfm, 25 psig discharge | 60 | CS | | BESSF-820 | 2 | $59,400 | $118,800 | Peters and Timmernaus 1980 |
| PB-817A/S | LP Vent Blower | 2400 cfm, 20 psig discharge | 60 | CS | | BESSF-830 | 2 | $73,000 | $146,000 | Peters and Timmernaus 1980 |
| GA-960 | Sterilization Tank Agitator | Single Impeller | 10 | SS 304 | | BESSF-960 | 1 | $10,800 | $10,800 | Chemcost |
| GA-961 | Cleaning Tank Agitator | Single Impeller | 10 | SS 304 | | BESSF-960 | 1 | $10,800 | $10,800 | Chemcost |
| GF-901 | Sand Filter | 34ft dia. X 8ft high | | CS | Tank | BESSF-920 | 1 | $39,600 | $39,600 | Icarus 1985 |
| GT-912 | Cooling Tower System | 54000 gpm | 110 | | 0.6 scaling factor | BESSF-940 | 1 | $751,400 | $751,400 | Badger 1984 |
| GU-903A/B | Demineralizers | 200 gpm | | CS | 0.6 scaling factor | BESSF-930 | 2 | $308,200 | $616,400 | Badger 1984 |
| GU-904A/S | Condensate Polisher | 1400 gpm | | SS | | BESSF-930 | 2 | $100,000 | $200,000 | |
| GU-907 | Hydrazine Addition Package | 150 gal tank, 2 pumps, 1 agitator | 1 | SS | | BESSF-930 | 1 | $15,000 | $15,000 | |
| GU-908 | Ammonia Addition Package | 150 gal tank, 2 pumps, 1 agitator | 1 | SS | | BESSF-930 | 1 | $15,000 | $15,000 | |
| GU-909 | Phosphate Addition Package | 150 gal tank, 2 pumps, 1 agitator | 1 | SS | | BESSF-930 | 1 | $15,000 | $15,000 | |
| GV-906 | Dearator | 1700 gpm, 17000 gal | | CS shell/ SS internals | | BESSF-930 | 1 | $133,000 | $133,000 | Badger 1984 |
| GY-910 | Instrument Air Dryer | 600 scfm/ desiccant | | CS | | BESSF-940 | 1 | $23,100 | $23,100 | Icarus 1985 |
| GZ-911 | Turbo Generator | 28 MW | | | | BESSF-940 | 1 | $6,500,000 | $6,500,000 | ABB Quote 1990 |
| HB-901A | Steam Boiler | 1100 psia, 450000lb/h, 300F superheat | | | $19,800,00 Installed | BESSF-910 | 1 | $0 | $0 | ABB Quote 1990 |
| PC-911 | Air Compressor | 1000 scfm | 250 | CS | 0.7 scaling factor | BESSF-940 | 1 | $58,300 | $58,300 | A.D. Little 1984 |
| PK-950A/B/S | Air Compressor Package | 28000 scfm | 2500 | CS | 0.7 scaling factor | BESSF-950 | 3 | $600,800 | $1,802,400 | A.D. Little 1984 |
| PK-951 | Chilled Water Package | 3380 gpm, 50 F | 2360 | CS | | BESSF-950 | 1 | $600,000 | $600,000 | |
| | | | | | | | Total Equipment Cost | | $21,634,900 | |
| | | | | | | | Total HP Required | | 16785 | |

Biomass To Ethanol
Simultaneous Saccharification & Fermentation
Project No. DOE/AFP/BESSF/02

EQUIPMENT LIST
TOWERS

| ITEM # | EQUIPMENT NAME | DIA/HGHT (feet) | NO. TRAYS | PRESS. psi | TEMP. F | MAT'L | REMARKS | DRW # | NO. REQ'D | PURCHASED COST/UNIT | TOTAL PURCHASED COST | SOURCE |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| AS-601 | Beer Column | 17.0/32 | 16 | 15 | 300 | CS | | BESSF-610 | 1 | $205,100 | $205,100 | Icarus 1985 |
| AS-602 | Rectification Column | 13.5/48 | 24 | 15 | 300 | CS | | BESSF-620 | 1 | $192,500 | $192,500 | Icarus 1985 |

Total Equip. Cost: $397,600